THE UNIVERSITY OF ALBERTA

MULTILOOP COMPUTER CONTROL OF AN EVAPORATOR

BY

© BRIAN A. JACOBSON

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The undersigned certify that they have read, and recommend to the Faculty of Graduate Studies for acceptance a thesis entitled "Multiloop Computer Control of an Evaporator" submitted by Brian A. Jacobson in partial fulfilment of the requirements for the degree of Master of Science.

ABSTRACT

The control of a pilot plant scale double effect evaporator, which is interfaced with an IBM 1800 Data Acquisition and Control System, was implemented using direct digital control and a generalized computer control program. The control program enabled the extension from single loop control to that of multiloop and multivariable control.

A series of 42 documented experimental runs were conducted on the double effect evaporator in order to evaluate the performance of the control systems developed.

The range of applicability of a linearized state space model of the evaporator, required for the advanced control techniques incorporated within the control program, was tested. It was found to satisfactorily predict the response of the primary controlled variables for disturbances with magnitudes of 20 percent about the point of linearization. The use of the control program, which was non-core resident, made it desirable to use a larger control interval so that the required computer time could be minimized. This lead to an investigation into the effect on the evaporator performance when larger sampling times were used in the primary control loops. Conventional DDC control was used and a significant deterioration in control was observed using a sampling time of two minutes and higher. In conjunction

with this study, a base case was established using DDC feedback control with a one minute sampling time and was used as a standard for comparing other control techniques.

The extension to multiloop and multivariable control was achieved using a generalized control law in matrix notation which incorporated feedback and feedforward control modes. Conventional integral control was implemented as well as a modified proportional plus integral control mode, which compensated for the effective dead time inherent in digital control by analytically predicting the controlled variables by a time equivalent to the deadtime. The predictor algorithm essentially eliminated deterioration of control when larger sampling times were used which resulted in satisfactory control at a sampling time as high as 8.0 minutes. An inferential feedback control system based on the state space model was also developed and gave control equivalent to that obtained using standard DDC control. Multivariable feedforward controllers were derived from the mathematical model in order to provide steady state compensation. The combined multiloop feedback and multivariable feedforward control modes resulted in a significant improvement in the overall control.

The use of the general control program to implement the above control systems demonstrates the ease at which advanced control techniques can be incorporated using direct digital control computers, and is the main advantage of computer control over conventional analog control.

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CHAPTER I

INTRODUCTION

1.1 Previous Work

Developments in process control have been illustrated by the control studies which have been implemented using the pilot plant scale, double effect evaporator at the University of Alberta. Andre (1) and Wilson (2) carried out feedback control studies using conventional analog instrumentation. Wilson also built a feedforward controller from basic analog components to compensate for measurable load disturbances entering the evaporator. Subsequent to these studies, the evaporator was relocated in a new building at which time it was fully interfaced to an IBM 1800 Data Acquisition and Control Computer. This allowed for data logging and direct digital control (DDC) of the double effect evaporator along with other process control applications and research projects. Fehr (3) developed a digital control system for the evaporator which used feedback and/or feedforward control modes. Inferential feedback control was also implemented using a simple mathematical model of the evaporator. This model was also used to derive feedforward controllers with and without dynamic compensation.

Using this work as a starting point, the current study demonstrates further the capabilities of the digital control computer in the process control field.

1.2 Objectives of Present Study

This control study provided an extension from the single loop control concept to that of multiloop and multivariable control. The double effect evaporator and IBM 1800 control computer, used in the previous control study (3), were used, however control was implemented using a combination of DDC and a general control program.

The objectives were to:

1. develop, using a generalized control program, a modular implementation for multivariable control for use in this study and by others in the department.
2. Implement multiloop and multivariable control techniques using the control program.

The multiloop/multivariable control techniques required a mathematical model of the process for implementation, therefore the linearized form of the five equation nonlinear model was used. This required establishing the range of applicability of the model in predicting the response of the evaporator.

The general control program was non-core resident and therefore it had to be transferred from bulk storage at each control interval (queued). In order to minimize the required computer time, it appeared desirable to use larger control intervals on the primary control loops (i.e. first effect liquid level, separator liquid level, and product concentration). Therefore direct digital feedback control was used to investigate the effect on the evaporator performance when

larger control intervals were used on the primary control loops. At this time a base case was also established in order to have a standard set of runs for comparing the multiloop and multivariable control techniques. Standard DDC proportional plus integral feedback control was used and the control interval was set so that no severe sampling time effects were observed.

Multiloop and multivariable control techniques were implemented using a generalized control law in matrix notation, and only the standard state space model formulation was required in order to apply these techniques to another process control application. The control law was to incorporate feedforward and feedback control algorithms, and therefore it was proposed to include the following algorithms:

1. standard proportional plus integral control
2. feedforward control
3. inferential feedback control
4. a modified proportional plus integral control mode which compensates for process dead times

in order to demonstrate multiloop feedback and multivariable feedforward control on the double effect evaporator using the control program.

A series of experimental runs were conducted in order to establish the range of applicability of the mathematical model, and to evaluate the multiloop/multivariable control techniques. The basic experimental equipment used is outlined in Chapter II while Chapter III devotes its discussion to the validity of the mathematical model of the

evaporator. Chapters IV and V were written to be essentially "stand alone" discussions of the development and final implementation of multivariable control. The overall conclusions are summarized in Chapter VI.

CHAPTER II

EXPERIMENTAL EQUIPMENT

2.1 Introduction

The double effect evaporator has undergone a number of physical changes since it was originally designed and built by Andre (1). Wilson (2) redesigned several sections in order to improve the operation. These included the recirculation in the second effect, the condensate collection system, the vacuum control, and the steam supply. After the termination of Wilson's experimental work, the evaporator was moved to a new building and rebuilt under his supervision. At that time, the system was redesigned in order to improve the evaporator performance and to enable continuous operation. Although the major equipment remained unchanged, a new feed system was added, additional piping was installed that provides for more than twenty different modes of operation, and new orifice plates and valve trims were added to allow for higher throughputs. From a control point of view, the most significant addition was the interfacing with the IBM 1800 Data Acquisition and Control Computer. Fehr (3) was the first to run the evaporator under direct digital control (DDC). He established standard operating conditions resulting in realistic loads on the evaporator which physically limited increased throughputs due to the separator capacity and pressure restriction on the glass.

Only minor changes have been made since the completion of Fehr's work, therefore only a brief description of the process and controls is given in this chapter. A simplified schematic of the evaporator, showing the forward feed mode of operation used in the previous and current study, is given in Figure 2.1. A more detailed flow diagram is available in reference (4).

2.2 General Description

The basic elements of the evaporator consists of two effects; a calandria-type and a long-tube vertical, forced circulation evaporator. The total feed flow (F8) to the first effect is made up of two streams, water and triethylene glycol solution, which have been preheated and mixed. In the forward feed mode steam (F1) also enters the first effect and is used to heat the liquid in this effect. The liquid flow from the first effect (F2) constitutes the feed for the second effect. This effect is run under forced circulation and is operated under a vacuum so that the steam produced in the first effect can be used to heat the second effect. The concentrated product from the second effect (F6) is cooled and recycled back to the solution feed tanks. The water condensate from the second effect steam chest (F5) and the overhead vapor from the separator (F7), which is condensed and cooled, are also recovered and returned to the water feed tanks.

A more detailed description of the evaporator can be found in references (1), (2), (3), and (4).

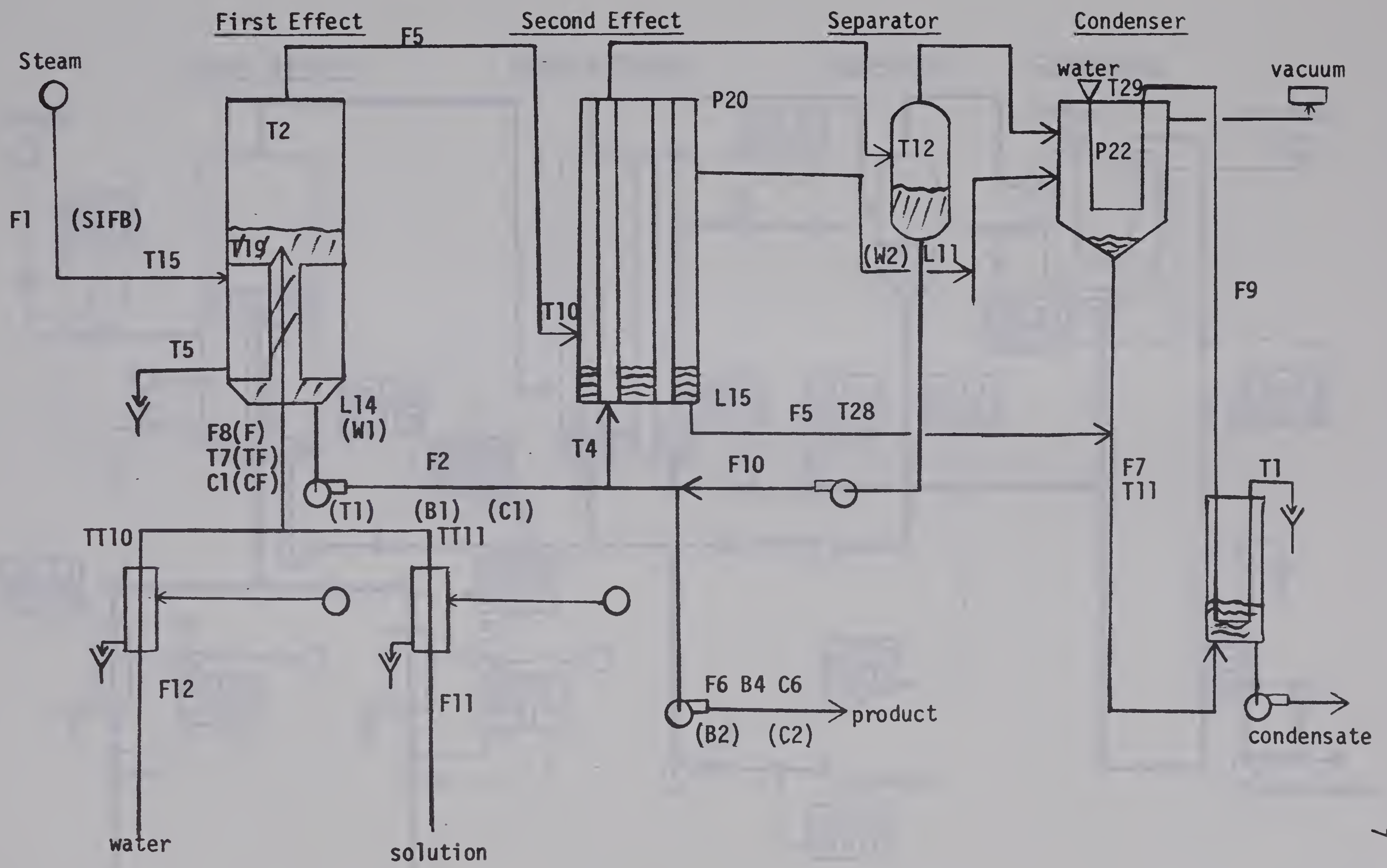


Figure 2.1 Double Effect Evaporator

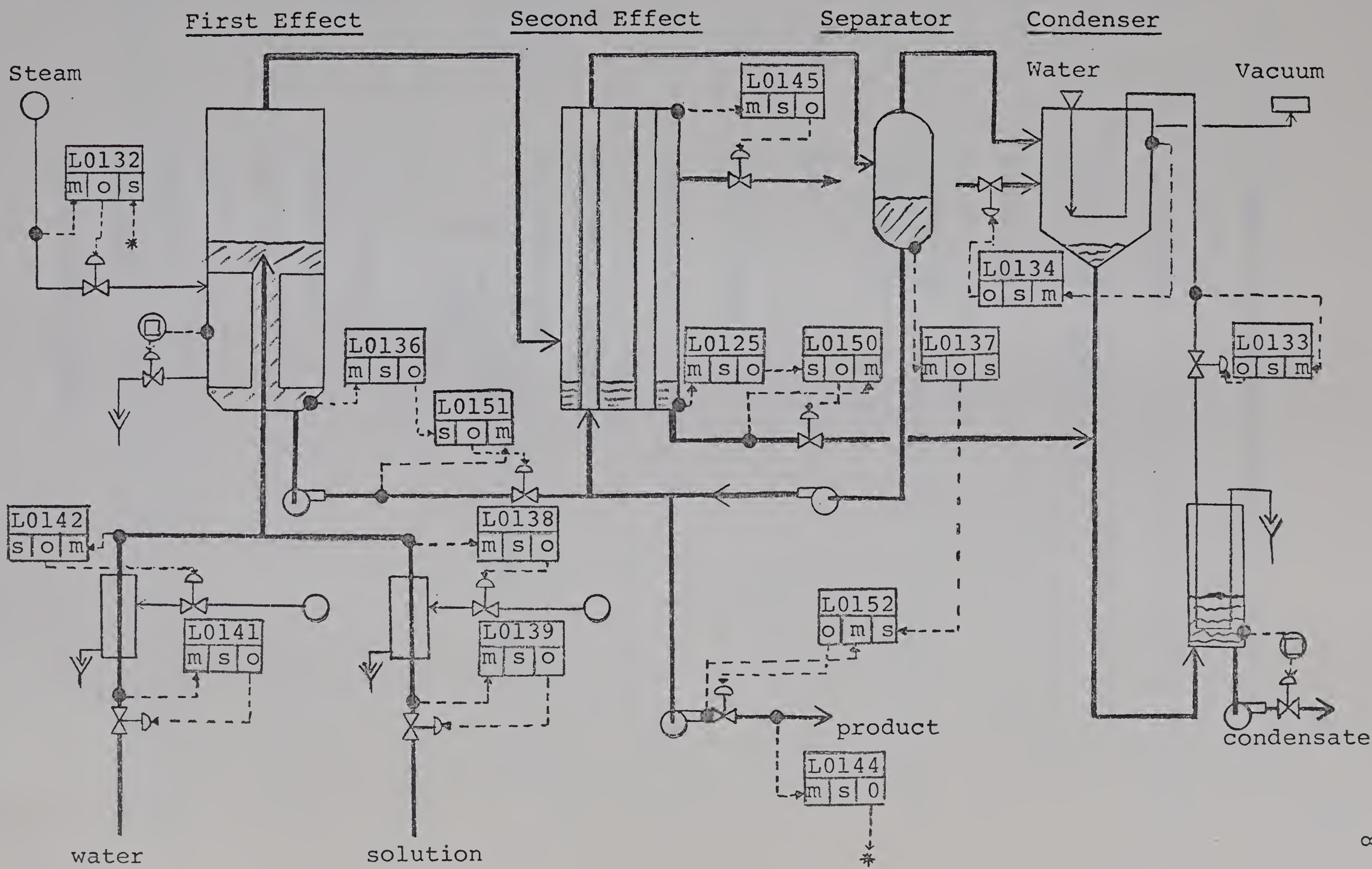
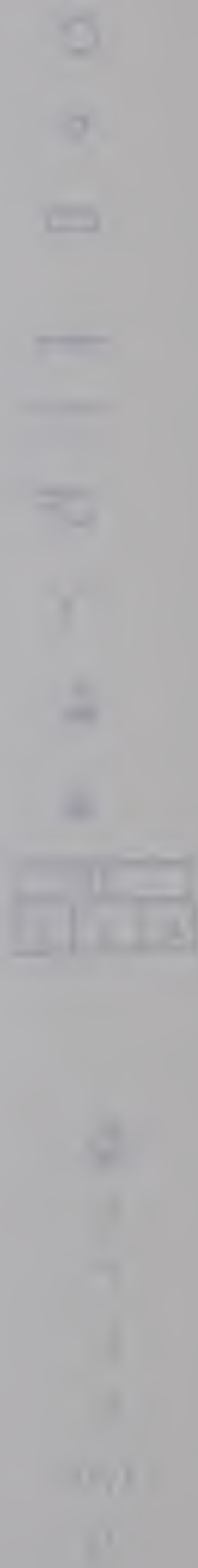


Figure 2.2 Evaporator Control Configuration

Legend for Figure 2.1 and 2.2

○	steam supply
▽	cooling water supply
▭	vacuum supply
━━	process line
——	utility line
	pump
Y	drain to sewer
	control valve
●	measurement transmitter
ID M O S	DDC Control Loop ID - control loop identification number M - measurement O - output S - setpoint
	analog controller
C	concentration
F	flow rate
L	level
P	pressure
T,TT	temperature
()	denotes model variable



2.3 Evaporator Controls

2.3.1 Control Configuration

Two different instrumentation systems are currently available for control of the double effect evaporator, conventional analog instrumentation and DDC control using the IBM 1800. During the experimental program, control was implemented using the IBM 1800. The basic DDC control configuration used is shown in Figure 2.2.

The control concept for the evaporator was such that the feed variables consisting of flow, temperature and concentration were not available for manipulation, which is typical of conditions in industry. In this study, disturbances were applied to the evaporator by introducing step changes in the feed variables. The control system for the evaporator feed consists of four controllers which control the flow and temperature of the two feed streams, water and triethylene glycol solution. The ratio control system used by Fehr was discarded since the inline refractometer, which was used to measure the feed concentration, was not functioning properly. In order to maintain the feed concentration at the desired level during an experimental run, solution was drawn from one of the solution feed tanks and product was returned to the other. Therefore one tank remained at a constant concentration throughout the run and the water to solution ratio could be set in order to produce the desired feed concentration.

The control of the various liquid levels was implemented using a combination of conventional analog and DDC controllers. The first ef-

fect level, the separator liquid level and the second effect condensate liquid level were controlled using the cascaded DDC control systems shown in Figure 2.2. Averaging liquid level control (5), which had been implemented by Fehr, was retained on the first effect and separator levels. The second effect condensate level was kept under "tight" control so that the liquid flow from the second effect steam space could be used as a measure of the first effect overhead vapor flow. The actual level transmitters were used to control the levels in the condensate system. The output from the transmitters were sent directly to the positioners on the control valves bypassing any analog controllers. With this system, only proportional control was available and the gain was changed by adjusting the spans on the transmitters.

The vacuum in the second effect and the cooling water flow to the condenser were both regulated using DDC feedback control loops. A vent controller on the second effect steam space was available to remove any air that accumulated, however, during the majority of the runs this controller was left on manual with the valve approximately three quarters open. Steam losses through the vent were monitored but found to be negligible.

The product concentration was controlled by manipulating the steam flow rate. An inline refractometer was used to obtain a continuous measurement of the product concentration which was used in the cascaded DDC control system. Feed forward control was also implemented, using conventional DDC control loops to compensate for feed flow disturbances.

As shown in Figure C-1, static feed forward control was implemented by effectively adjusting the steam flow rate when changes in the first effect bottoms flow were detected. Fehr showed that this feed forward system was equivalent to using dynamic feed forward control based on the feed flow to the evaporator.

The basic feedback control system presently being used has evolved from experience over the past few years but can also be derived from a consideration of the state space model of the evaporator (6). A more detailed development can be found in references (1), (2), and (3).

2.3.2 Control Problems

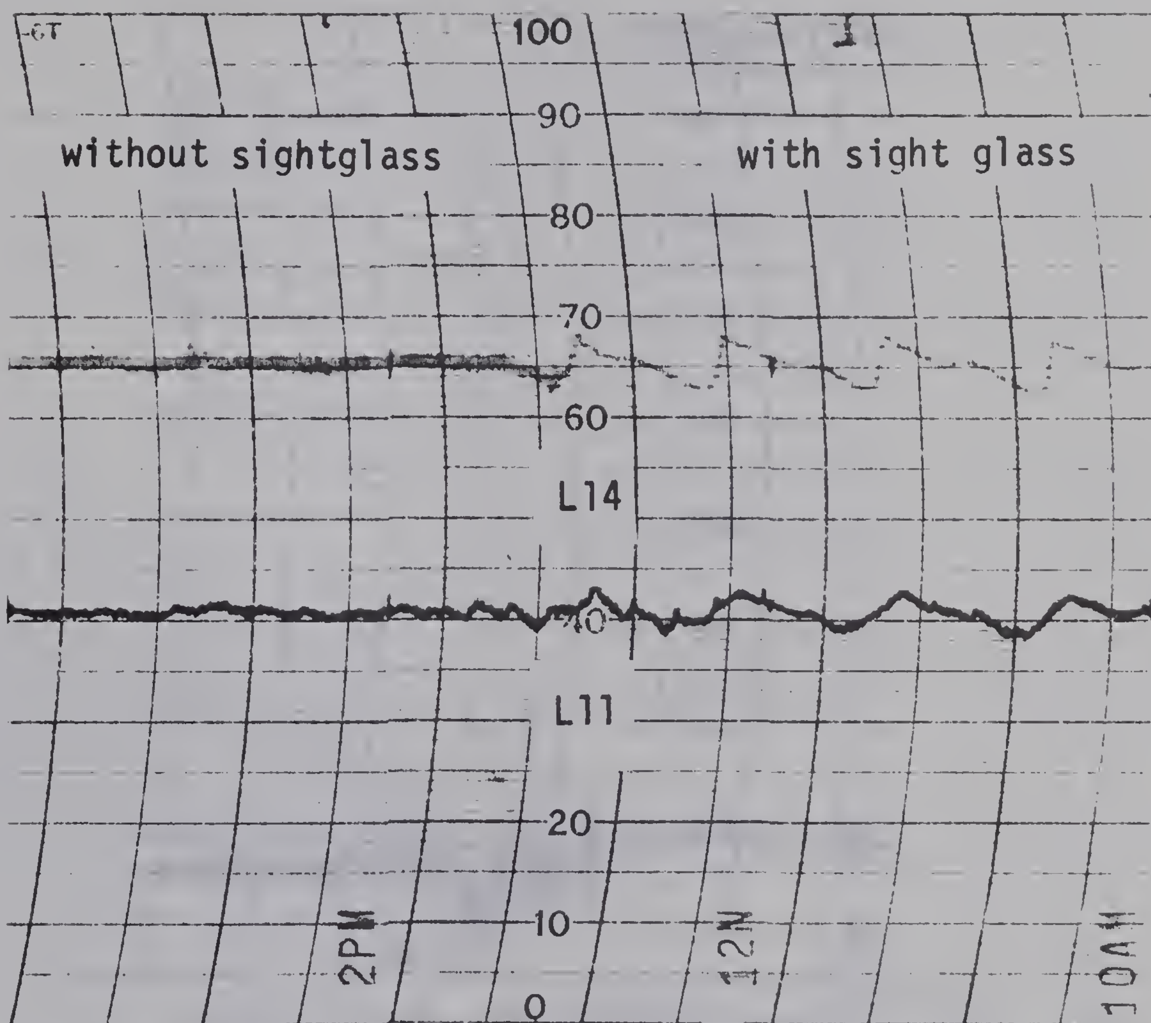
A number of control problems were encountered during the experimental program, the most significant of which are presented below. Although these did not usually present any serious difficulty during an experimental run, their presence hindered the continuous operation of the evaporator.

Early in the experimental program, problems were encountered with the indicated first effect liquid level. The indicated liquid level would rise slowly for about thirty minutes and then would drop by 1 - $1\frac{1}{2}$ inches. This phenomenon was observed by Fehr who attributed it to water vapor condensing in the copper tubing between the first effect vapor space and the sight glass causing a pressure differential between the two. When this pressure difference reached a certain value, the water column was blown out and the indicated level reading would fall as shown

in Figure 2.3. In order to alleviate this problem, the tubing to the sight glass was reconfigurated avoiding any horizontal runs, the tube size was increased to 3/8 inch, and both the reference and measurement legs from the d-p cell were connected to the sight glass. However, the problem persisted and was finally eliminated by connecting the measurement leg from the d-p cell directly to the first effect rather than the sight glass. This resulted in a noisier reading, Figure 2.3, however, the cycling in level was eliminated.

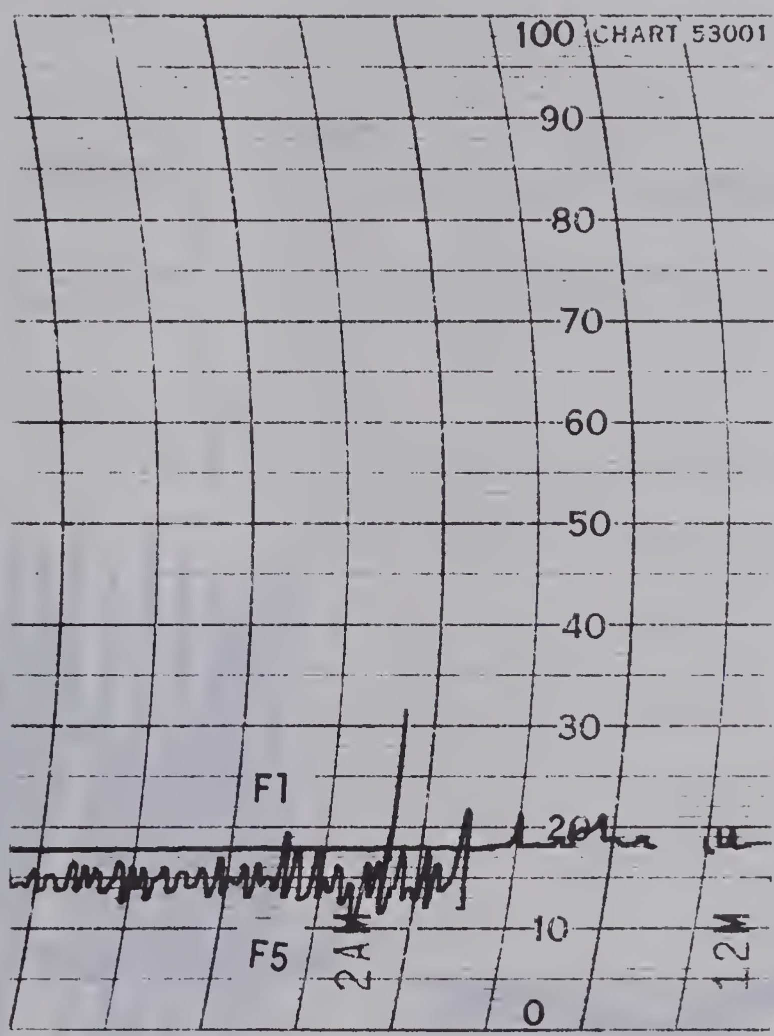
Condensate from the first effect steam chest was removed by allowing it to run into a small glass vessel on the outlet of the chest which replaced a conventional steam trap. The liquid level in the glass vessel was controlled using a d-p cell transmitter and control valve to give smooth "proportional" action. Occasionally the liquid level in the sight glass would be lost, due to vaporization of the water in the transmitter's reference leg, allowing the steam to pass directly to the drain sewer which caused a "spike" in the flow, Figure 2.4. The control action would close the valve to compensate for the larger flow resulting in the drop below the setpoint. This effect was eliminated by maintaining a very small water flow to the reference leg.

The condensate flow from the condenser to the condensate tank produced a noisy flow reading as shown in Figure 2.5. Since this flow was used in the overall mass balance, large errors of closure were experienced. It appeared that air was being pulled through the orifice due to the pressure differential between the condenser and condensate



L14 span = 31.5 in. of water
L11 span = 25.0 in. of water

Figure 2.3 Recording of the First Effect Liquid Level With and Without the Sight Glass



F1 span = 4.2 lbs./min (√ chart reading)

Figure 2.4 Recording of Flow Spike in the Steam Flow

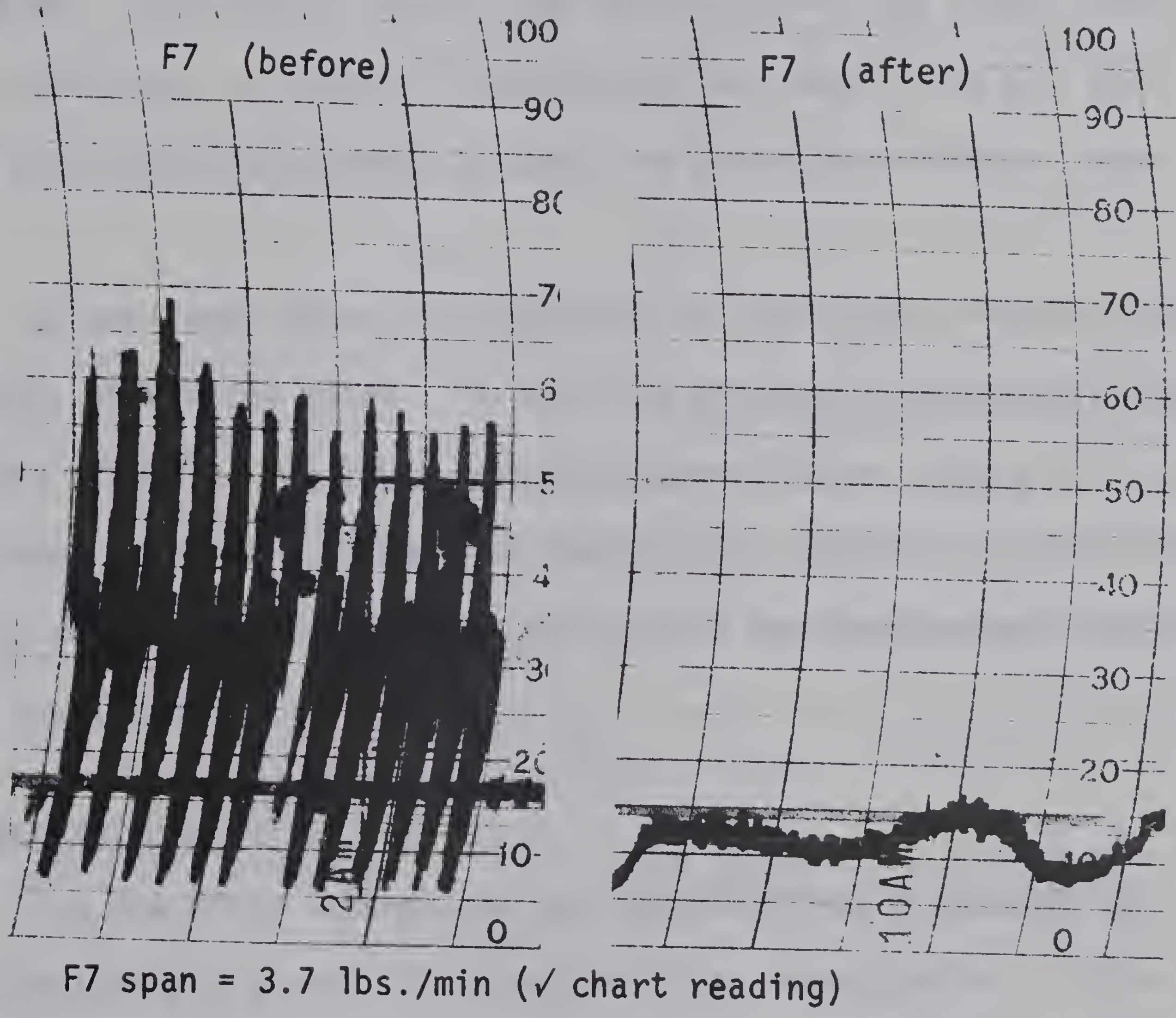


Figure 2.5 Recordings of Separator Overhead Flow With and Without a High Noise Level

tank which are both under vacuum. The pressure difference resulted from fluctuations in the vacuum supply pressure which were more pronounced in the condensate tank. An improved reading was observed, Figure 2.5, after the piping on the vacuum system was changed by putting a damping tank between the vacuum supply and the two vessels, and by relocating the orifice to the discharge of the rundown tank condensate pump. Tight level control was maintained on the liquid level in the rundown tank so that any accumulation was negligible and the measured flow could be assumed to equal the separator overhead vapor flow.

As mentioned above, fluctuations in the vacuum pressure were experienced, at various times. To show the effect of these small changes (0.3 inches of Hg) on the primary controlled variables Figure 2.6 has been included. In order to maintain much closer control on this variable the piping was changed as outlined above and a new measurement transmitter was also installed.

2.4 Evaporator Operation

The operating concept for the double effect evaporator has changed considerably during the course of this investigation. Prior to the installation of the IBM 1800, the evaporator was run as a batch process since an operator was not available twenty four hours each day to supervise the operation. Since the IBM 1800 is available continuously a program was written to monitor the evaporator operation when an operator was not there. Running in this type of standby mode enables experimental

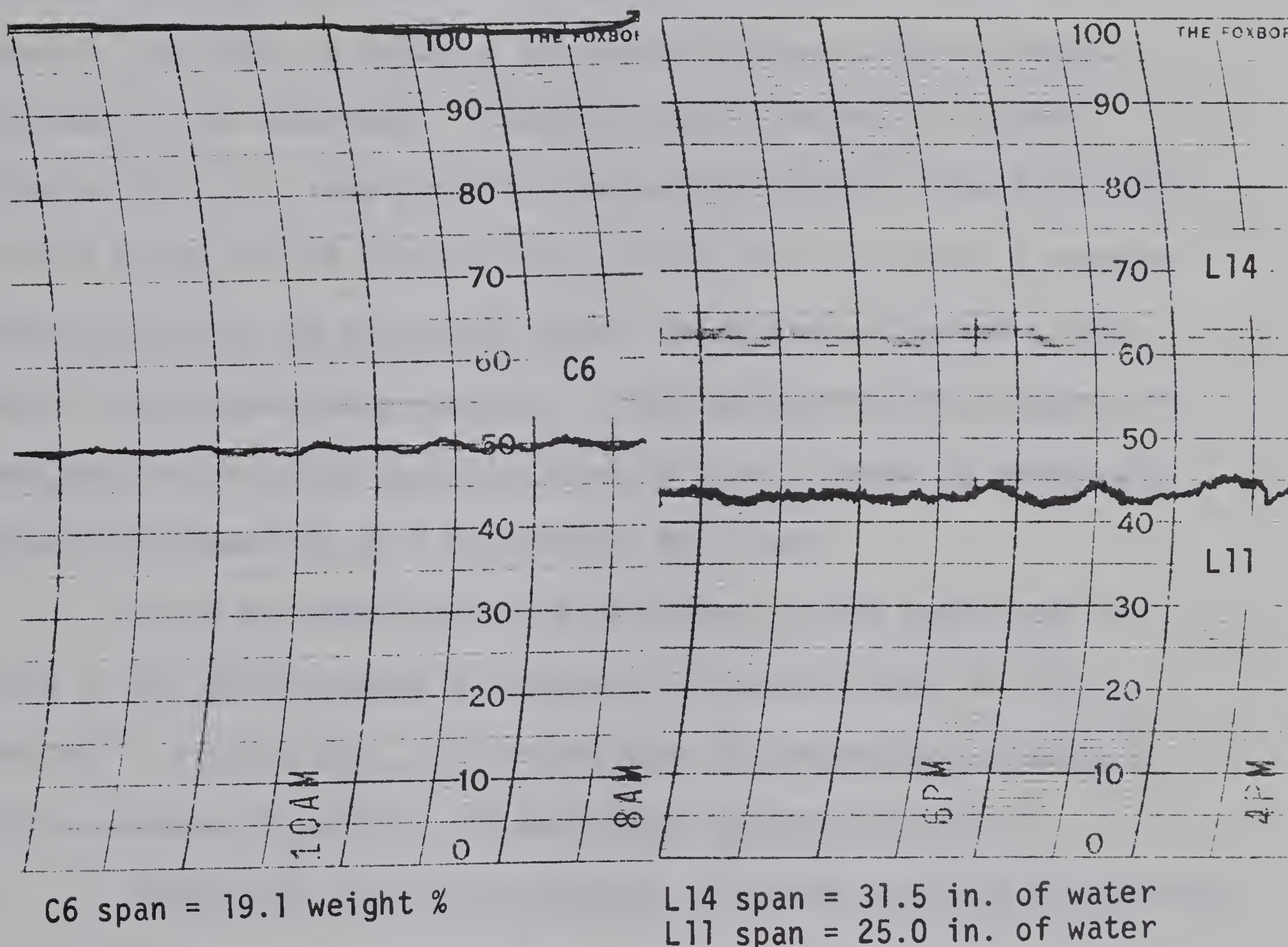


Figure 2.6 Recordings of First Effect Level, Separator Level and Product Concentration

runs to be made at any time.

A monitor program was written, which is run on the IBM 1800 at a preset time interval, to check key variables on the evaporator against a set tolerance limit. When a particular variable exceeds the corresponding tolerance limit, the evaporator is shutdown by the computer. In order to ensure a fail safe shutdown certain changes were made to the evaporator. These included changing all control valves to fail to a safe position, installing normally closed electric solenoid valves on the utility supply lines, and installing a magnetic breaker switch on the main power supply which when activated by the computer shuts down the evaporator. Also, only the minimum number of pumps required for this operation were left on in order to reduce the chance of shutdown due to a malfunction of a pump.

Since the completion of this project, a new option was included in DDC which enables a program to be executed when an error is detected in any DDC loop. This eliminates the necessity of having a program executed to monitor the particular process application.

Along with the monitor program, an effort was made to formalize the operating procedures for the evaporator. Although this was not part of the research project, it was felt that these procedures should be better defined since graduate and undergraduate students who are unfamiliar with the evaporator would be required to use it for research, course work and/or term projects. A set of procedures were written (Appendix D) which include:

1. start up procedure
2. experimental run procedure
3. continuous operation procedure
4. a procedure for switching between local analog control and DDC.

These procedures were supplemented by a program which eliminates the need of the user to have a knowledge of the mechanical operations which are required when operating under DDC control.

2.5 Experimental Run Documentation

Each of the experimental runs were documented using a standard procedure. The steady state data listed in the tables of Appendix A were obtained from three Fortran programs:

1. a program which collected the major flows, pressures, temperatures and concentrations (7).
2. a program to read all the temperature readings which have been interfaced to the IBM 1800 (Appendix D, 24).
3. a program which calculates the errors of closure in the mass, component and energy balances around the evaporator and has the capability of data adjustment using a linear programming technique (8).

The output of these programs, the Foxboro recording charts, and any comments about the experimental runs were kept in a data book which is available in the department. The transient data was plotted using a

"Calcomp" plotter (eg. Figure A-1a) which is interfaced with the IBM 1800. The data for the plotting routines was obtained from punched cards which have been retained for future reference. The variables marked on the graphs are the model nomenclature which is available from Chapter III and Figure 2.1

CHAPTER III

MATHEMATICAL MODEL

3.1 Introduction

Use of a process control computer permits more sophisticated control than is possible using conventional instrumentation. However, to implement most of these advanced control techniques, a mathematical description of the process is required. It is the purpose of this chapter to outline the development of the model used in this project and also to describe the experiments which demonstrate the range of applicability of the model.

3.2 Model Development

The pilot plant scale, double effect evaporator can be represented by the simplified schematic drawing in Figure 2.1. The main dynamic elements, the first and second effect liquid holdups, can be considered as perfectly mixed or lumped parameter systems in order to derive a simplified mathematical model. Also, the dynamics of the first and second effect steam chests and the transportation lags of the process piping can be neglected (1). Using these assumptions, Andre (1) derived a mathematical model by writing the mass, component and energy balances around each effect. This resulted in a set of six nonlinear differential equations plus a set of algebraic equations. These were further reduced to five nonlinear differential equations by neglecting

the energy accumulation in the second effect since the pressure (temperature) in this effect is closely controlled.

The nonlinear model was linearized, with the variables in normalized perturbation form, resulting in a set of linear differential equations (9) which can be expressed in state space notation by:

$$\dot{\underline{X}} = \underline{\underline{A}}\ \underline{X} + \underline{\underline{B}}\ \underline{U} + \underline{\underline{C}}\ \underline{D} \tag{3.1a}$$

$$\underline{Y} = \underline{\underline{E}}\ \underline{X} \tag{3.1b}$$

where $\underline{X}$, $\underline{U}$, $\underline{D}$, and $\underline{Y}$ are the state, manipulated, disturbance and output vectors, respectively and $\underline{\underline{A}}$, $\underline{\underline{B}}$, $\underline{\underline{C}}$ and $\underline{\underline{E}}$ are the time invariant coefficient matrices. The solution of equation (3-1) using standard techniques (10) was obtained

$$\underline{X}(n+1) = \underline{\underline{\Phi}}\ \underline{X}(n) + \underline{\underline{H1}}\ U(n) + \underline{\underline{H2}}\ \underline{D}(n) \tag{3.2}$$

using a program (11) which calculates the state difference equation matrices. This program assumes $\underline{U}$ and $\underline{D}$ are constant over the time interval specified and the solution is exact if this assumption holds. The corresponding difference equation with a one minute time basis is shown in Table 3.1. The coefficient matrices of equation (3.1) were calculated using the reference steady state given in Table A-1.

The model equations (Table 3.1) are used to calculate the

Table 3.1

<u>Five Equation Linearized Model</u>

$$\begin{bmatrix} W1(n) \\ C1(n) \\ h1(n) \\ W2(n) \\ C2(n) \end{bmatrix} = \begin{bmatrix} 1.0 & -0.0008 & -0.1043 & 0.0 & 0.0 \\ 0.0 & 0.8960 & 0.0978 & 0.0 & 0.0 \\ 0.0 & -0.0047 & 0.3450 & 0.0 & 0.0 \\ 0.0 & -0.0007 & -0.0936 & 1.0 & 0.0001 \\ 0.0 & 0.0440 & 0.0938 & 0.0 & 0.9544 \end{bmatrix} \begin{bmatrix} W1(n-1) \\ C1(n-1) \\ h1(n-1) \\ W2(n-1) \\ C2(n-1) \end{bmatrix} + \begin{bmatrix} -0.1107 & 0.0 & -0.0182 \\ 0.0 & 0.0 & 0.0175 \\ 0.0 & 0.0 & 0.1828 \\ 0.0890 & -0.0468 & -0.0163 \\ -0.0412 & 0.0 & 0.0163 \end{bmatrix} \begin{bmatrix} B1(n-1) \\ B2(n-1) \\ SIFB(n-1) \end{bmatrix}$$

$$+ \begin{bmatrix} 0.1648 & -0.0 & -0.0085 \\ -0.0546 & 0.1049 & 0.0082 \\ -0.0169 & -0.0003 & 0.0856 \\ 0.0015 & -0.0 & -0.0076 \\ -0.0028 & 0.0025 & 0.0076 \end{bmatrix} \begin{bmatrix} F(n-1) \\ CF(n-1) \\ hF(n-1) \end{bmatrix}$$

$$\begin{bmatrix} W1(n) \\ W2(n) \\ C2(n) \end{bmatrix} = \begin{bmatrix} 1.0 & 0.0 & 0.0 & 0.0 & 0.0 \\ 0.0 & 0.0 & 0.0 & 1.0 & 0.0 \\ 0.0 & 0.0 & 0.0 & 0.0 & 1.0 \end{bmatrix} \begin{bmatrix} W1(n) \\ C1(n) \\ h1(n) \\ W2(n) \\ C2(n) \end{bmatrix}$$

current state vector $[W1\ C1\ h1\ W2\ C2]^T$ or the output vector $[W1\ W2\ C2]^T$ from a knowledge of the previous state, the measured manipulated vector $[B1\ B2\ SIFB]^T$, and the measured disturbance vector $[F\ CF\ TF]^T$. If the previous state vector, $\underline{X}(n-1)$, is actually the process measurements at time (n-1) then the model is only used to predict over one time interval. However, if the previously calculated state is used, then the calculated state and output vectors at time (n) represent the response of the model for changes in $\underline{U}$ and $\underline{D}$ over n time intervals. This latter method is used in this chapter in order to compare the response of the actual process to those calculated using the mathematical model.

3.3 Experimental Verification

3.3.1 Experimental Objectives

The purpose of the first part of the experimental program was to establish the range of applicability of the five equation linearized model in predicting the open-loop response of the double effect evaporator. This was accomplished by comparing the process and model responses, for step disturbances in the feed flow, feed temperature, feed concentration or steam flow, with no control on the product concentration. Although the state variables, with the exception of the first effect product concentration, can be continuously measured; only a comparison between the actual and calculated output variables (W1, W2, C2) were made. These variables constitute the primary controlled variables on the double effect evaporator (1, 2, 6) and will be the major concern in the subsequent control study.

Concurrent with this project, other work in the department (9, 12) was being done which required open loop data from the evaporator. These included an evaluation of various theoretical evaporator models as well as a method for empirically fitting experimental data to a given model form. This resulted in a further incentive to obtain a complete set of open loop data.

3.3.2 Discussion of the Open Loop Results

The experimental runs which were conducted have been summarized in Table 3.3. The magnitude of the various disturbances, which were used to force the evaporator and the mathematical model, are relative to their respective reference states shown in Table 3.2a. The control parameters used in the primary control loops during the open loop runs are given in Table 3.2b.

Early in the experimental program, a significant deviation between the responses of the actual and calculated liquid holdups (levels), W1 and W2, was observed. This was attributed to model errors and the integrating nature of the holdup terms in the model equations (i.e. corresponding diagonal elements in the Φ matrix are 1.0 or in matrix $\underline{\underline{A}}$ the diagonal elements are zero for those variables). The difference equation solution is exact if $\underline{U}$ and $\underline{D}$ are constant over the time interval which for this case was one minute. However, since the process measurements of $\underline{U}$ and $\underline{D}$ were used to force the model, the above assumption is not valid since the process response curves are continuous rather than discrete, and therefore change during the model time interval. Since the levels calculated by the model are a function of their input

Table 3.2a

Process Reference Condition

Process Variable Description	Model Variable	Reference Value
Total Feed Flow	F	5.0 lbs./min.
Total Feed Temperature	TF	190.0 degrees F
Total Feed Concentration	CF	0.030 weight fraction*
Steam Flow	SIFB	2.0 lbs./min.

*triethylene glycol in water

Table 3.2b

Feedback Controller Reference Condition

Control Loop Description	Model Variable	Poll Time (seconds)	Proportional Constant	Integral Constant (seconds)
First Effect Liquid Level	W1	2	1.0	2048.0
Separator Liquid Level	W2	2	3.0	2048.0
Product Concentration	C2	Open loop -- on manual control		

Table 3.3

Open Loop Runs

Description of Disturbance	Run Number	Magnitude of Disturbance in Percent*
Feed Concentration	OL8	-30.0
Feed Flow Rate	OL9	+10.0
Feed Flow Rate	OL10	-20.0
Feed Flow Rate	OL11	+20.0
Feed Flow Rate	OL12	-10.0
Feed Temperature	OL13	-17.9
Steam Flow Rate	OL18	+10.0
Steam Flow Rate	OL15	-20.0
Steam Flow Rate	OL16	+20.0
Steam Flow Rate	OL19	-10.0

$$*\% \text{ Change} = \frac{\text{Change in Disturbance}}{\text{Process Reference Value}}\ 100.0$$

Note: Changes shown as 20.0% are from $\mp$ 10.0% to $\pm$ 10.0% with respect to the reference process values

and output flows, any inaccuracies due to the continuous nature of these variables or any measurement noise encountered in discrete sampling results in a steadily increasing deviation when "integrated" over an extended time period. To alleviate this problem, the linear model was modified to include the two level controllers (Appendix B) resulting in a seventh order model which was used in all the subsequent work to calculate the responses of the output variables (Table 3.4).

A steady state offset resulted between the actual product concentration as measured on a bench refractometer, and that indicated by the product inline refractometer in a number of the experimental runs. Since the offline analysis was more accurate than the inline refractometer, the product concentration data from the inline refractometer were linearly adjusted to agree with the offline data at the final steady state to avoid erroneous conclusions from being made. This procedure was validated when it was discovered that the discrepancy was due to range changes on the inline refractometer caused by scaling (film deposit) on the prism and/or temperature changes of the electronics associated with the inline part of the refractometer. An attempt is now being made to completely eliminate this problem which persisted to some extent throughout the research project.

The mathematical model was linearized at the standard reference state given in Table 3.1a and Table A-1. The disturbances, with the exception of feed temperature and concentration, were introduced so that they were always within 10 percent of the corresponding reference state

Table 3.4

Seven Equation Linearized Model

$$\begin{bmatrix} W1(n) \\ C1(n) \\ h1(n) \\ W2(n) \\ C2(n) \\ X6(n) \\ X7(n) \end{bmatrix} = \begin{bmatrix} .9130 & -.0009 & -.1097 & .0 & .0 & -.0013 & .0 \\ .0 & .8666 & .1062 & .0 & .0 & .0 & .0 \\ .0 & -.0049 & .2822 & .0 & .0 & .0 & .0 \\ .0621 & -.0008 & -.0969 & .8676 & .0001 & .0009 & -.0019 \\ -.0313 & .0526 & .1031 & .0 & .9445 & -.0005 & .0 \\ .9559 & -.0006 & -.0675 & .0 & .0 & .9993 & .0 \\ .0322 & -.0005 & -.0592 & .9324 & .0 & .0005 & .999 \end{bmatrix} \begin{bmatrix} W1(n-1) \\ C1(n-1) \\ h1(n-1) \\ W2(n-1) \\ C2(n-1) \\ X6(n-1) \\ X7(n-1) \end{bmatrix} + \begin{bmatrix} -.0266 \\ .0261 \\ .2239 \\ -.0233 \\ .0242 \\ -.0098 \\ -.0086 \end{bmatrix} \begin{bmatrix} SIFB(n-1) \end{bmatrix}$$

$$+ \begin{bmatrix} .2102 & -.0001 & -.0118 \\ -.0706 & .1343 & .0116 \\ -.0234 & -.0004 & .0996 \\ .0095 & -.0001 & -.0104 \\ -.0080 & .0039 & .0108 \\ .1063 & -0.0 & -.0044 \\ .0033 & -0.0 & -.0038 \end{bmatrix} \begin{bmatrix} F(n-1) \\ CF(n-1) \\ hF(n-1) \end{bmatrix}$$

$$\begin{bmatrix} W1(n) \\ W2(n) \\ C2(n) \end{bmatrix} = \begin{bmatrix} 1.0 & 0.0 & 0.0 & 0.0 & 0.0 & 0.0 & 0.0 \\ 0.0 & 0.0 & 0.0 & 1.0 & 0.0 & 0.0 & 0.0 \\ 0.0 & 0.0 & 0.0 & 0.0 & 1.0 & 0.0 & 0.0 \end{bmatrix} \begin{bmatrix} W1(n) \\ C1(n) \\ h1(n) \\ W2(n) \\ C2(n) \\ X6(n) \\ X7(n) \end{bmatrix}$$

as shown in Figure 3.1. Therefore, in the worst case the model is

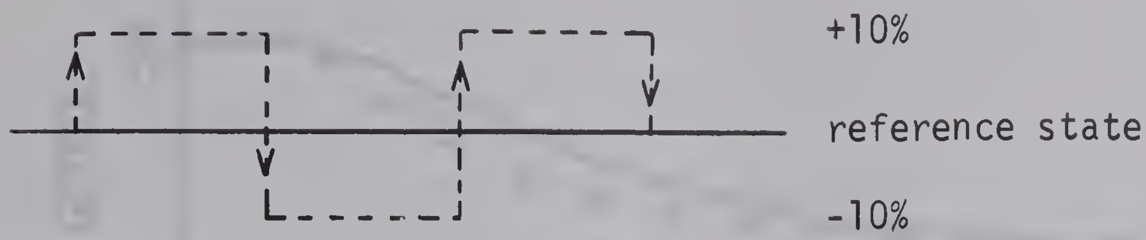


Figure 3.1 Disturbance Profile

basing its calculation on a disturbance with a magnitude of 20 percent about the point of linearization. For the experimental runs which did not start at the process reference state, the model was forced to this nonreference state before the actual $\underline{U}$ and $\underline{D}$ data were introduced into the model. The model transients were shifted so that the process and model output variables would start at the same point, which permitted direct comparison of the two transient responses.

The response curves of the output variables for a decrease in feed concentration are shown in Figure 3.2. The C2 model response tends to lead the process, which was intuitively expected since the linear model contains no transportation lags and was based on perfect mixing, which is not the case in the actual process. There is good agreement at the final steady state, considering the 30 percent step away from the point of linearization, indicating the process is relatively linear for this type of disturbance. A similar phenomenon was observed when the process was forced using feed temperature, Figure A-6a, however the final steady state offset on C2 was more significant. The actual and calculated levels changed very little on the concentration forced run, which is also indicative of the temperature forced and $\pm$ 10 percent

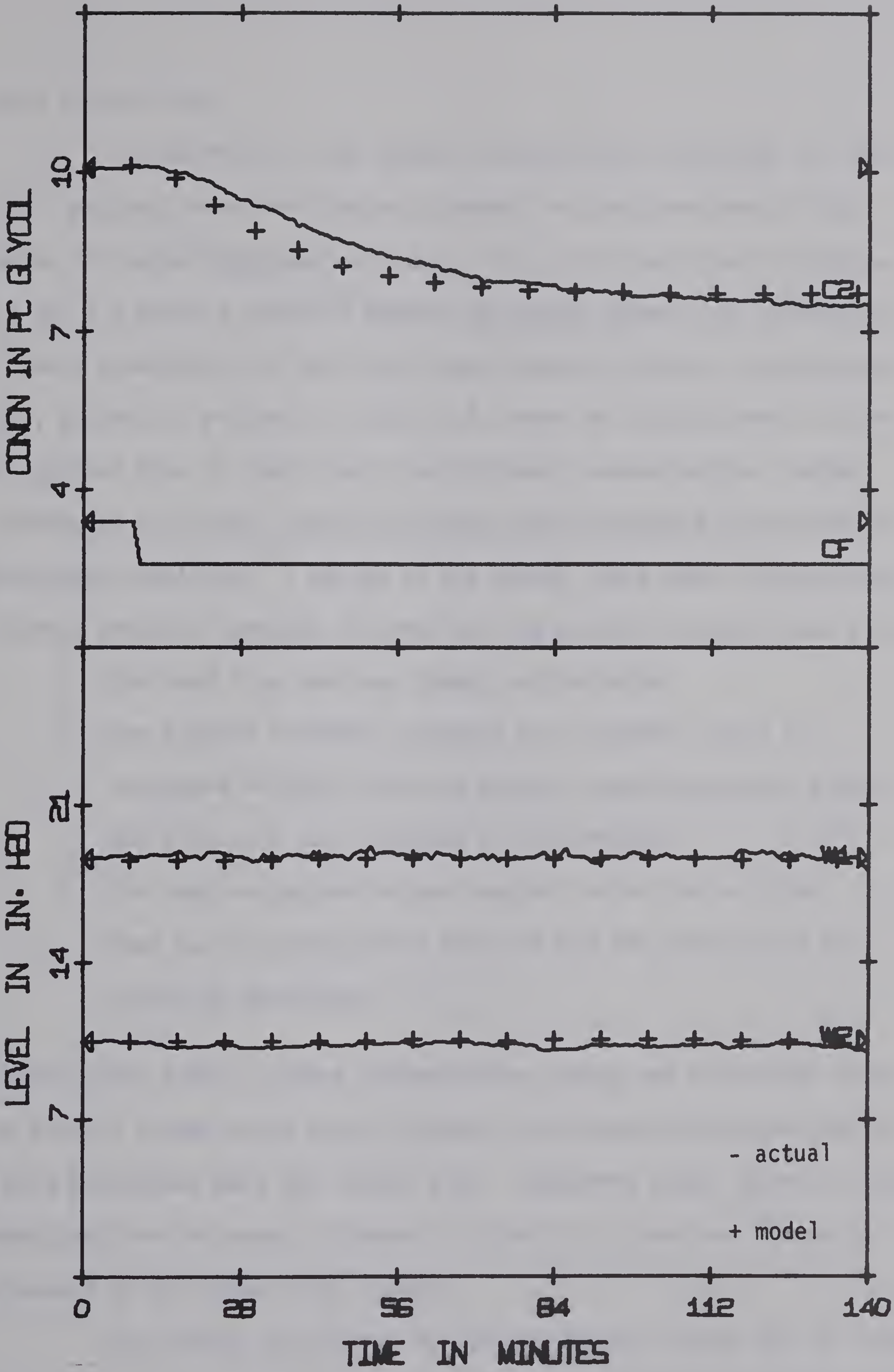


Figure 3.2 Response of the Output Variables for Run OL8 (-30% in CF under DDC feedback control with no control on C2)

steam forced runs.

A comparison of the product concentration responses for the feed flow runs shows that better agreement was obtained when disturbances of larger magnitude were used. This is illustrated in Figures 3.3 and 3.4 where a positive step of 10 and 20 percent was introduced. A closer examination of the feed forced measured product concentration data, column $|\Delta C \text{ process}|$ in Table 3.5, shows an inconsistency for an up and down step in feed flow since different concentration changes (difference at initial and final steady state) resulted using steps of equivalent magnitude. A review of the steady state data indicated that the most probable variable in error was the product concentration since:

1. the feed flow rate was steady and reliable
2. the product flowrate, although not constant, could be estimated reliably from the Foxboro recording charts and/or the transient data recorded by the computer
3. the feed concentration was constant since the solution feed was only taken from one tank and the product was recycled to the other.

Therefore, the actual process concentration change was calculated from the overall steady state glycol balance relationship which resulted in a more consistent data set (Table 3.5). Comparing these values to those predicted from the model ($|\Delta C\text{model}|$ in Table 3.5) confirms the better agreement on the larger step changes.

The smaller deviations for the 20 percent versus the 10 percent

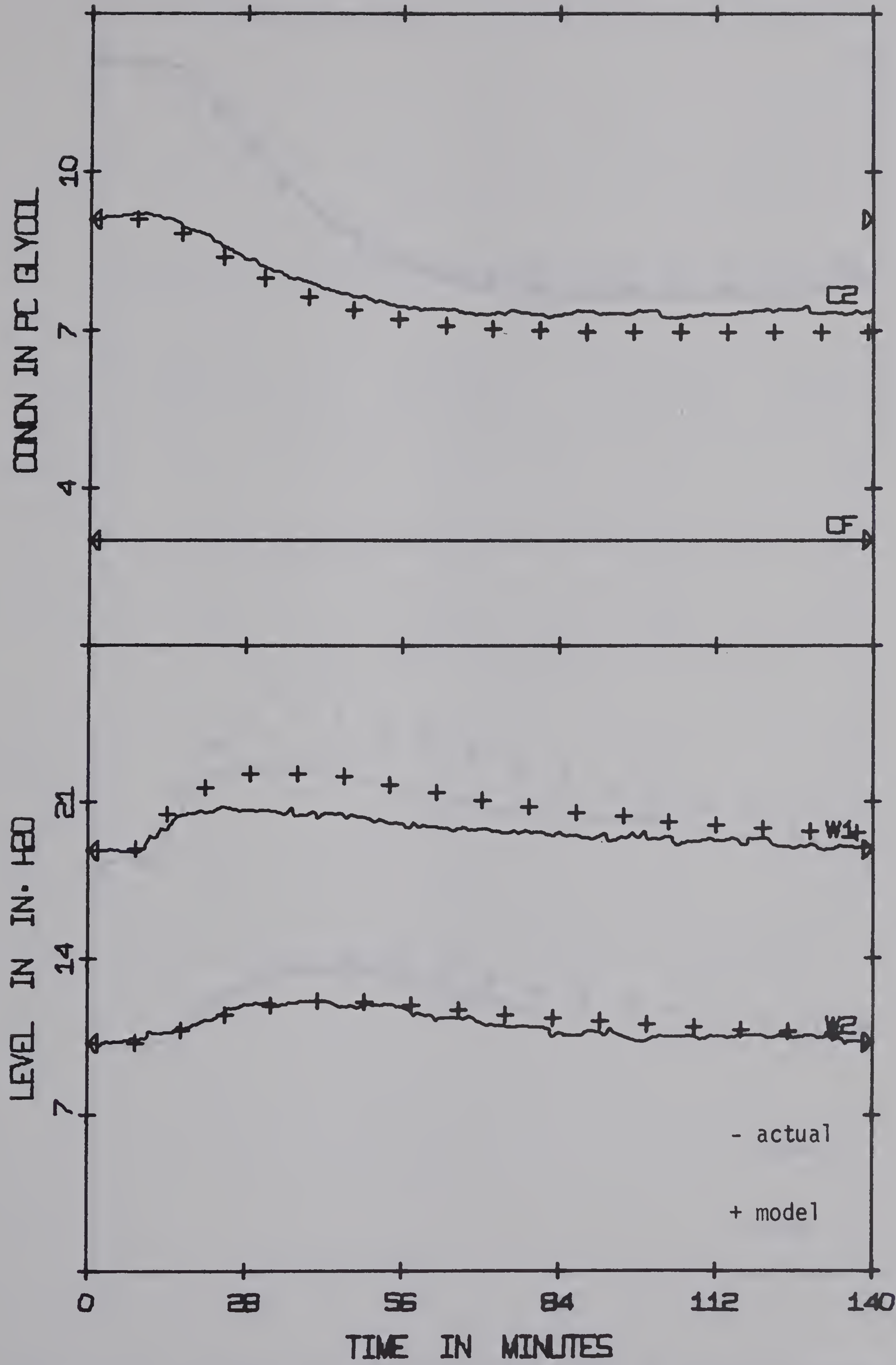


Figure 3.3 Response of the Output Variables for Run OL9 (+10% step in F under DDC feedback control with no control on C2)

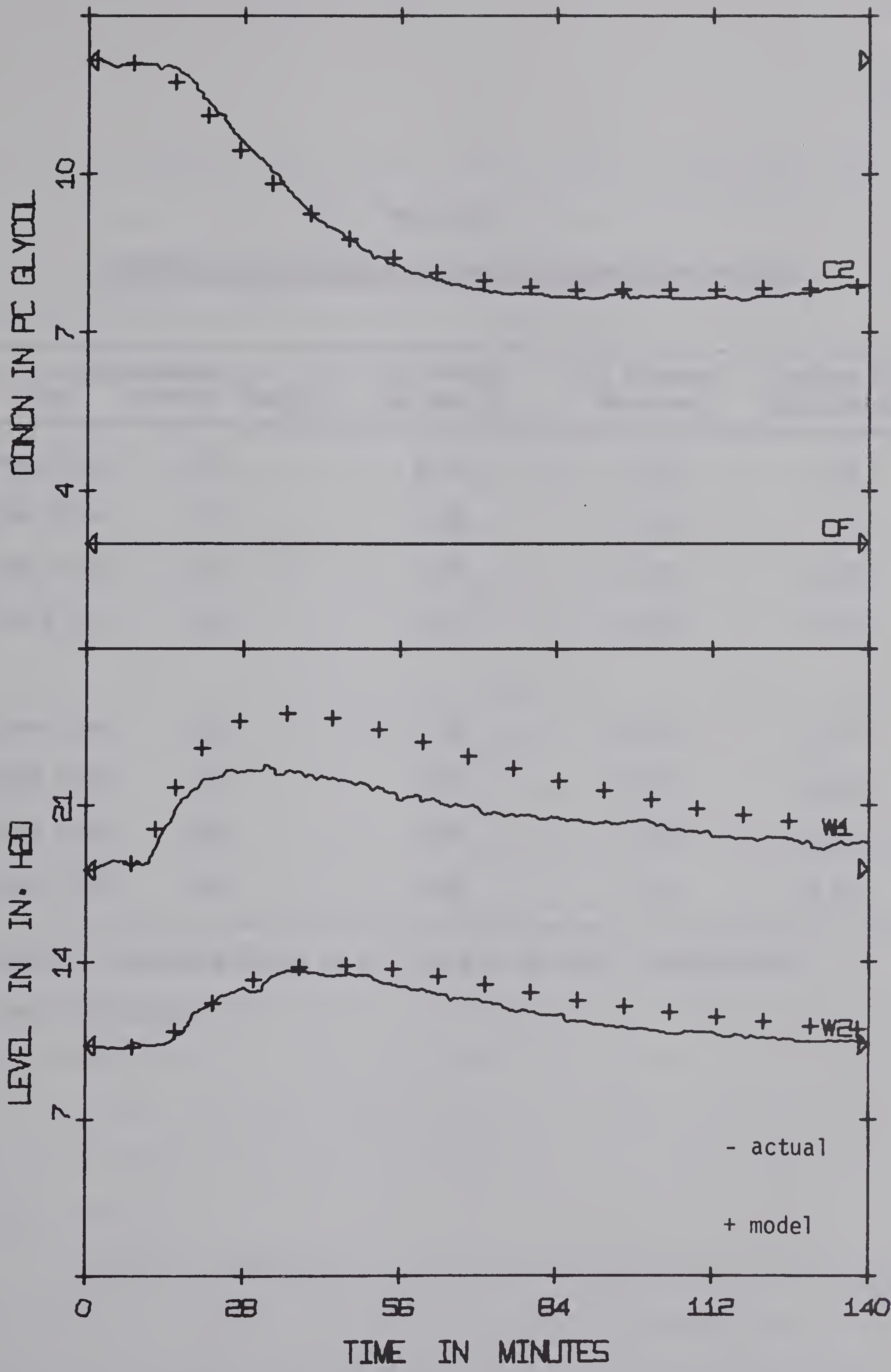


Figure 3.4 Response of the Output Variables for Run OL11 (+20% step in F under DDC feedback control with no control on C2)

Table 3.5

Concentration Changes for Feed and Steam Flow Forcing

Disturbance Type	Percent Change*	$\lvert\Delta C$ Model$\rvert$ (weight %)	$\lvert\Delta C$ Process$\rvert$ Measured	(weight %) Calculated
Feed Flow	+10	2.10	1.70	1.85
Feed Flow	-10	2.25	1.40	1.70
Feed Flow	+20	4.35	4.10	4.30
Feed Flow	-20	4.15	4.60	4.30
Steam Flow	+10	1.75	2.10	1.80
Steam Flow	-10	1.80	1.90	2.00
Steam Flow	+20	3.45	3.30	3.30
Steam Flow	-20	3.40	3.10	3.20

*Based on reference state of 5.0 and 2.0 lbs./min. respectively.
See Table 3.2a.

runs can be explained using Figure 3.5 which shows the nonlinear and linear C2-F gain curves. For a + 10% step change in feed flow, the final steady state product concentration calculated from the linear model is shown at point 5 in Figure 3.5 and there is a significant deviation from the actual process value (point 4). Prior to the +20% step in feed flow, the model was initialized to the nonreference initial steady state (point 2) and the calculated product concentration was adjusted to agree with the process (point 3). This adjustment effectively shifted the linearized gain curve resulting in the line 3-4 in Figure 3.5. When the model was then "forced", by changing the feed flow from 4.5 to 5.5 lb/min., the corresponding steady value of C2 changed along the line 3-4 and ended up at a value almost identical to the actual process concentration (point 4). Thus the final steady state predicted by the model was closer to the correct value after a 20% change (centered around the point of linearization) than it was for a 10% change (away from the reference state). The procedure of forcing agreement at the initial steady state between the process and model variables is particularly important when control is first initiated. It also facilitates comparison of the actual and model responses since in composite graphs such as Figure 3.3 the model and process response start with exactly the same initial conditions.

A similar adjustment was done on the steam forced concentration data, Table 3.5, however, the deviations between the adjusted and actual process product concentration changes were generally smaller than those

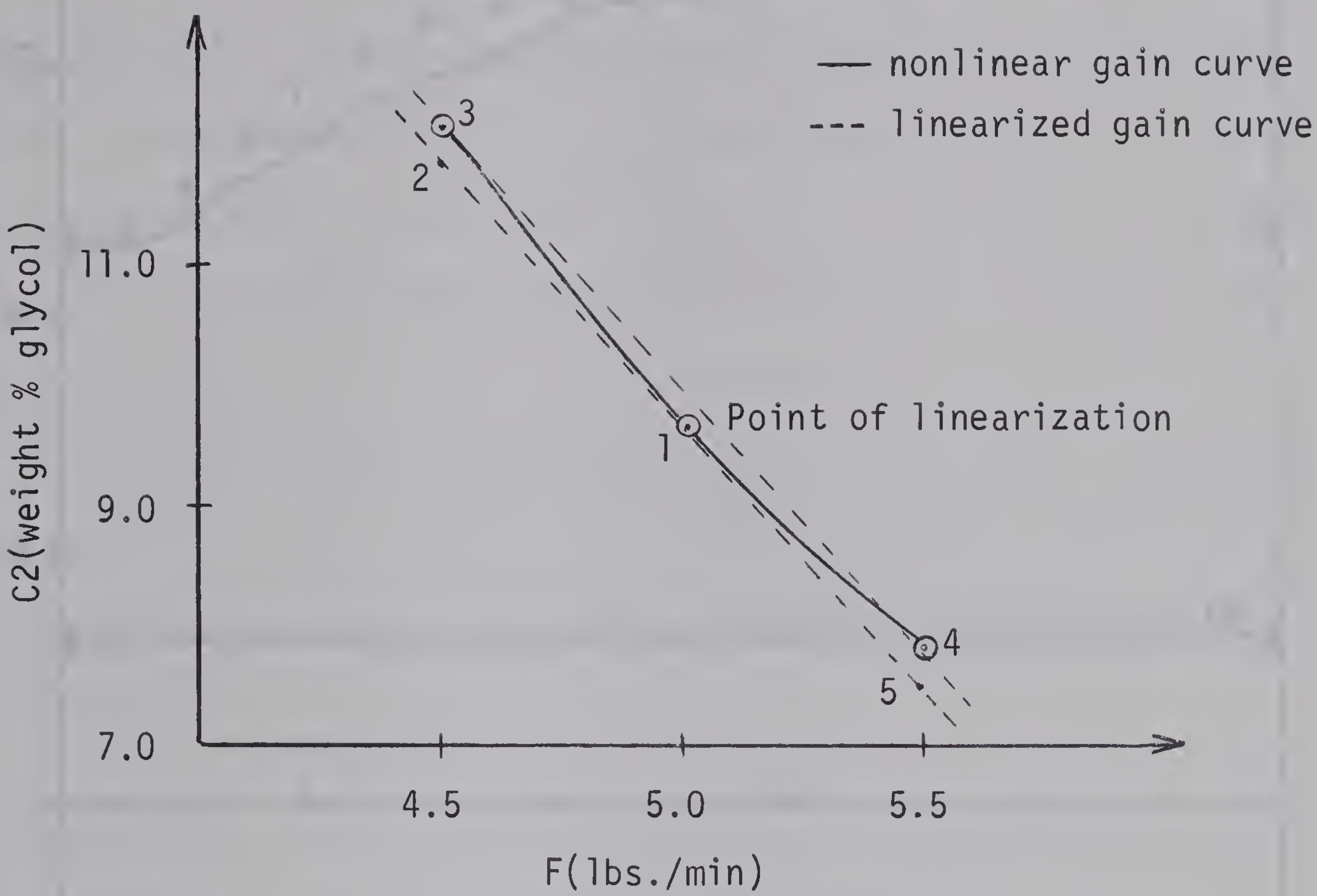


Figure 3.5 Linear and Nonlinear C2-F Gains

found for the feed forcing runs. The agreement for all the steam forced runs was good, as can be seen from the + 20 percent steam forced run in Figure 3.6, since the process C2-SIFB gain is relatively linear for the range of flows used. In both the feed and steam forced runs, the model response tends to lead the process C2 response.

The response of the first effect level (W1) predicted from the model deviates from the actual process for most of the transient in the feed forced and ± 20 percent steam forced runs (see Figures 3.3, 3.4 and 3.6). It appears to be due to a conversion factor which was used in converting from holdup (in pounds of solution) to the actual process measurement (inches of water). This conversion is complicated by the boiling liquid and changes in the cross sectional area with height

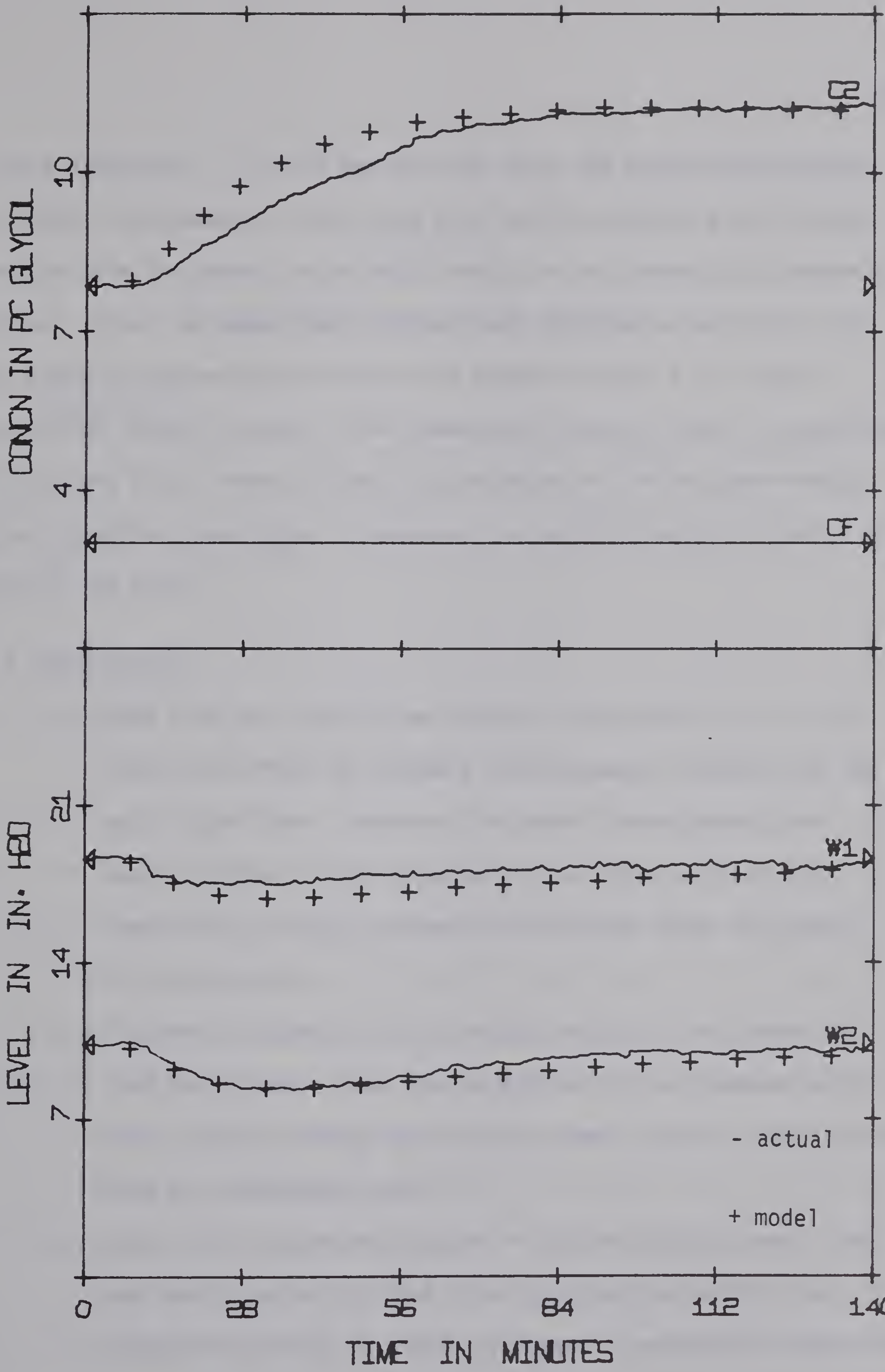


Figure 3.6 Response of the Output Variables for Run OL16 (+20% step in SIFB under DDC feedback control with no control on C2)

in the evaporator. If this was not the case, one would have expected a similar disagreement in W2 since both level controllers are incorporated within the model, which would result in an interaction between W1 and W2. Also, the model went further away from the setpoint for both the positive and negative steps which suggests that it is simply a conversion factor in error. The conversion factors, used in converting both W1 and W2 to inches of water, were based on limited experimental data; therefore, more data is necessary in order to obtain a better estimate of the gains.

3.4 Conclusions

1. Feed flow was found to be the most severe forcing variable since its effect on the mass and component balances was the most significant. However, the model showed good agreement relative to the response of the actual product concentration for step changes of 20 percent about the point of linearization.
2. The model response of the product concentration tended to lead the process since the assumptions of no transportation lags, perfect mixing, and omitting small dynamic contributions, were not completely valid.
3. Due to the integrating nature of the two liquid levels, the two level controllers had to be incorporated within the mathematical model in order to obtain a reasonable comparison between the process and model responses.

CHAPTER IV

FEEDFORWARD AND FEEDBACK CONTROL

4.1 Introduction

The availability of a process control computer permits advanced control techniques and methods to be implemented which utilize the computer's flexibility. It was the purpose of this work to initiate development for implementation of multiloop control, and ultimately the more complex multivariable control techniques, using a combination of a general control program and direct digital control (DDC).

The general control program, which was used to implement the multiloop and multivariable control, was non core resident and was brought into core (queued) every control interval and executed. The frequency of execution of any program (core or non core resident) directly affects the available computer time, and therefore determines the load capabilities of the computing system. This is particularly noticeable with a non core resident program since transfer from bulk storage is required, which uses proportionally more computer time.

This made it desirable to sample the control loops within the program as infrequently as dynamics permit so that the computer execution time is minimized (13). Therefore, an investigation into the effect on the evaporator performance, when larger sampling times were used in the primary control loops, was conducted using DDC feedback control.

A base case was also established using DDC feedback control and was used as the standard for purposes of comparing other control techniques. The primary control loops in the base case used conventional proportional plus integral control at a sampling time of 64 seconds.

Inferential feedback control was implemented by using the fifth order state space model (9) to calculate the controlled (output) variables, which were used in the control law in place of the measured variables. The control law was formulated using matrix (multiloop/multivariable) notation. Multivariable feedforward control was also developed to compensate for the measurable load disturbances and utilized the same mathematical model of the evaporator.

The base case, feedback control, and the feedforward plus feedback control studies are discussed in the following sections.

4.2 Theory

The mathematical model for the double effect evaporator was developed by writing mass, component and energy balances around each of the two effects and was linearized (9) about the process reference state (Table A-1). The solution for the linear state space time invariant model was obtained from a computer program (11) resulting in a set of difference equations which can be expressed in matrix notation by:

$$\underline{X}(n) = \underline{\underline{\Phi}}\,\underline{X}(n-1) + \underline{\underline{H1}}\,\underline{U}(n-1) + \underline{\underline{H2}}\,\underline{D}(n-1) \qquad (4.1a)$$

$$\underline{Y}(n) = \underline{\underline{E}}\ \underline{X}(n) \tag{4.1b}$$

where $\underline{X}$ is the state vector, $[W1\ C1\ h1\ W2\ C2]^T$

$\underline{U}$ is the manipulated vector, $[B1\ B2\ SIFB]^T$

$\underline{D}$ is the disturbance vector, $[F\ CF\ hF]^T$

$\underline{Y}$ is the output vector, $[W1\ W2\ C2]^T$

This model formed the basis for the inferential and feedforward multi-loop/multivariable control studies. Numerical values for the coefficient matrices are given in Table 3.1.

The use of proportional, integral and feedforward control modes, using multivariable techniques, results in a control law which in matrix-vector notation can be expressed by:

$$\underline{U}(n) = \underline{\underline{KP}}(\underline{Y}(n)-\underline{SP}(n)) + \sum_{i=0}^{n} \underline{\underline{KI}}(\underline{Y}(i)-\underline{SP}(i)) + \underline{\underline{KFF}}\ \underline{D}(n) \tag{4.2}$$

where $\underline{\underline{KP}}$, $\underline{\underline{KI}}$ and $\underline{\underline{KFF}}$ are the proportional, integral and feedforward control matrices, respectively. Multiloop control, which is a special case of multivariable control, results when the control matrices are diagonal.

Inferential control consists of controlling a process variable whose value is not measured, but inferred (3). For this study, inferential control was implemented using the mathematical model of equation 4.1 with the following modification:

$$\underline{XC}(n) = \underline{\underline{\Phi}}\ \underline{XS}(n-1) + \underline{\underline{H1}}\ \underline{U}(n-1) + \underline{\underline{H2}}\ \underline{D}(n-1) \tag{4.3a}$$

$$\underline{XS}(n) = \underline{XC}(n) + \underline{\underline{AX}}\ (\underline{XM}(n) - \underline{XC}(n)) \tag{4.3b}$$

$$\underline{Y}(n) = \underline{\underline{E}}\ \underline{XS}(n) \tag{4.3c}$$

where for inferential control $\underline{\underline{AX}}$ is a diagonal matrix given by

$$\underline{\underline{AX}} = \text{diag}[0.0\ 0.0\ 0.0\ 0.0\ 0.0] \tag{4.4}$$

Therefore, the calculated output variables ($\underline{Y}$) for use in the control calculations (equation 4.2) are based on the calculated state and the measured manipulated and disturbance vectors. The use of this technique enabled switching easily between inferential and conventional control (for conventional control $\underline{\underline{AX}}$ = diag(1.0 1.0 1.0 1.0 1.0)) as well as providing an extension to measurement filtering which in this form is a simplification of Kalman Filtering(14). This technique was used in part in subsequent work to predict the current value of an unmeasured process variable and it also provided a "backup" in case of instrument failure.

The multivariable feedforward controllers used in this work were derived from the mathematical model based on steady state compensation for load disturbances. From equation (4.1a) the feedforward contribution compensates for measurable load disturbances at steady state if:

$$\underline{\underline{H1}}\ \underline{UFF}(n) + \underline{\underline{H2}}\ \underline{D}(n) = \underline{0} \tag{4.5}$$

For the case where the dimension of the state and manipulated variables are equivalent, the feedforward control matrix can be calculated directly using:

$$\underline{\underline{KFF}} = - \underline{\underline{H1}}^{-1}\ \underline{\underline{H2}} \tag{4.6}$$

If the dimension of the manipulated variables is less than the state, equation (4.6) is not valid since the inverse of $\underline{\underline{H1}}$ does not exist. However, if feedforward control is formulated with the objective of minimizing the weighted sum of the steady state deviation, then it follows directly from a least squares analysis that

$$\underline{\underline{KFF}} = - (\underline{\underline{H1}}^T\ \underline{\underline{Q}}\ \underline{\underline{H1}})^{-1}\ \underline{\underline{H1}}^T\ \underline{\underline{Q}}\ \underline{\underline{H2}} \tag{4.7}$$

where $\underline{\underline{Q}}$ is a square weighting matrix (15). Equation (4.7) reduces to (4.6) when $\underline{\underline{H1}}$ is square.

4.3 Discussion of Results

For the majority of the experimental runs, feed flow was used as the disturbance variable and was forced symmetrically about the process reference state of 5.0 lbs./min. The steps were made from $\pm$ 10 percent to $\mp$ 10 percent relative to the reference steady state. This was judged a realistic load since feed flow disturbances have the most

significant effect on the evaporator and the magnitude is at the limit of applicability when using the linearized model (equation 4.1).

4.3.1 Base Case

As a starting point in establishing the base case for this research project, two runs were made under DDC feedback control using the control constants of Fehr (3) in the primary control loops (W1, W2, C2). The basic shape of the C2 response curves for a feed flow change of $\pm$ 20 percent, Figure 4.1, were similar to Fehr's results, however, the current runs show a faster rise from the setpoint and return to the setpoint in a shorter period of time. This difference was attributed in part to the product inline refractometer which has been cleaned and recalibrated since Fehr's experimental program resulting in a change in the overall loop gain. Therefore in order to have a comparison to Fehr's DDC feedback work, the above runs will be used.

The controller constants in the product concentration control loop were then increased in order to reduce the magnitude of the transient observed in Figure 4.1. The final values used were tested for feed flow disturbances of $\pm$ 10 percent and $\pm$ 20 percent, runs DDC1-3 and DDC6-7 respectively. For the smaller step size only a small deviation from the setpoint resulted (Figures A-11 to 13), however, the use of the larger steps (Figures A-16 to 17) had more of an effect on the system. This demonstrates another reason for using the larger steps since a more valid comparison between various runs can be made.

The above runs were made using relatively small sampling times

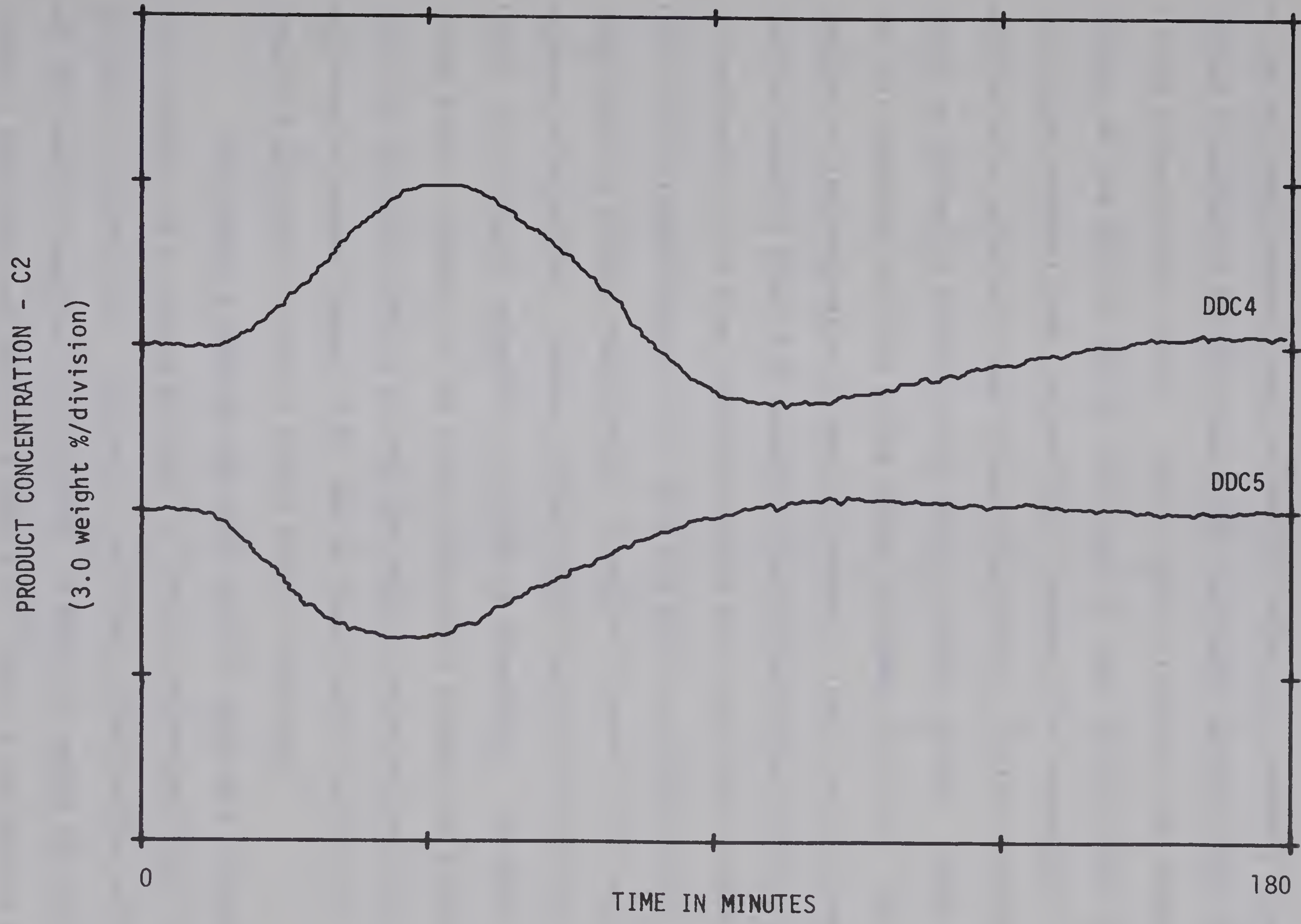


Figure 4.1 Response of the Product Concentration for ± 20% steps in Feed Flow (DDC feedback control)

on the two level control loops, W1 and W2 (Table 4.3), however, it was desired to establish the base case using reasonable sampling times on all the primary control loops. Digital control intervals are frequently selected on the basis of design rules such as "less than one tenth of the systems dominant time constant"(16). However, the evaporator is a multivariable system with some strong interactions, and time constants cannot be defined as exactly as for single variable systems. Based on the "63% response" in C2 to a step change in steam, the overall time constant of the evaporator is about 30 minutes. Thus it would appear that the sampling time of 64 seconds on the concentration control loop was reasonable. The time constants for the two liquid levels calculated from holdup/throughput (i.e. W1/F and W2/B1) are approximately seven and nine minutes respectively. A control interval of 64 seconds appears to be marginal, however, since "tight" control of the liquid levels is not usually essential, it might be acceptable. The determining factor, however, is the strong interaction between the bottoms flow from the first effect, B1, and the product concentration C2 (6). Any large or sudden changes in B1, due to level control on W1, will affect C2 and hence control of liquid level is important.

The sampling times on the level control loops were increased to 64 seconds and a step up and down in flow was introduced (Runs DDC8 and DDC9 in Figure 4.2). This had a significant effect on the control and the control loops had to be retuned in order to reduce the oscillations, Runs DDC10 and DDC14 in Figure 4.2. Since it was desired to

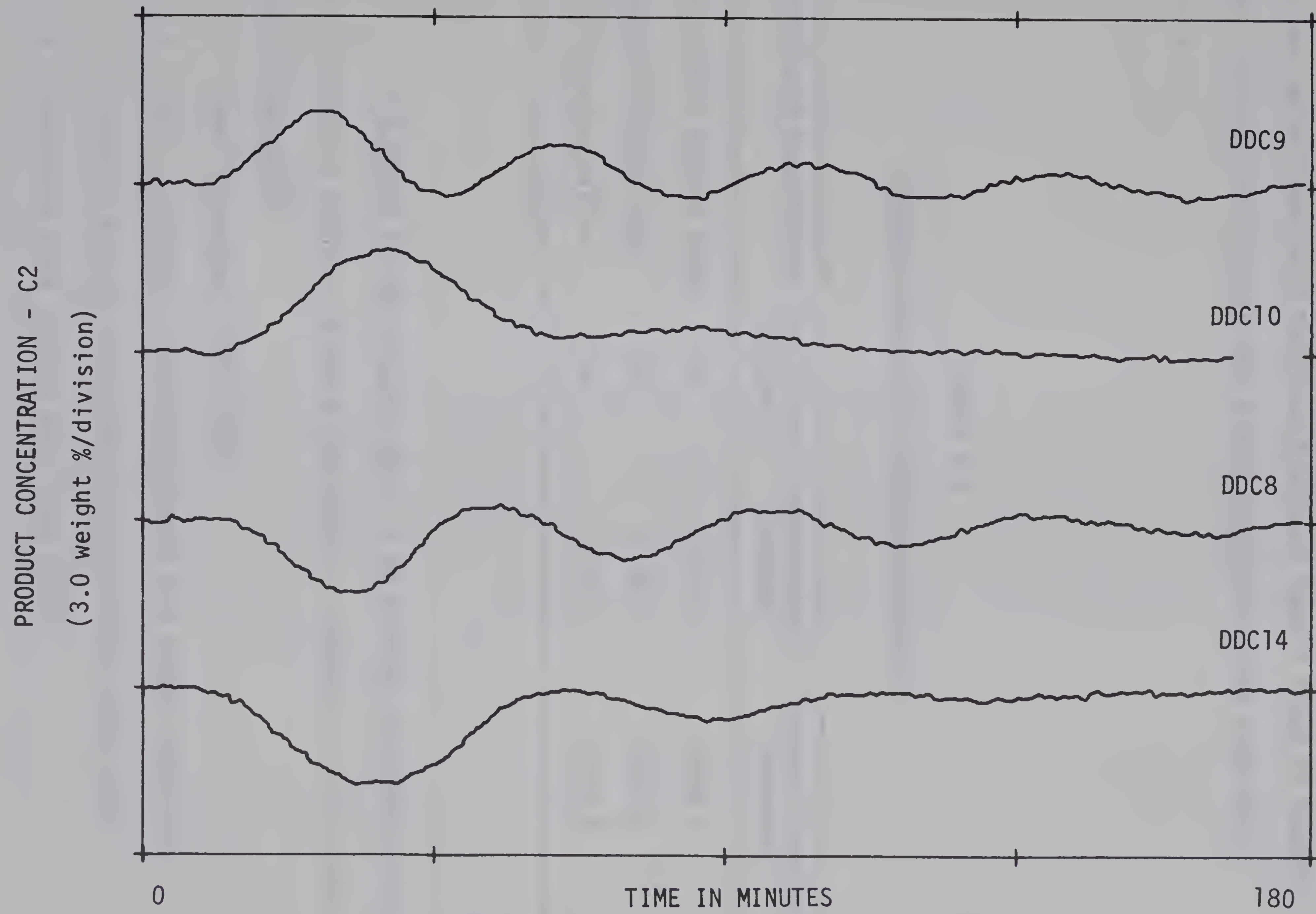


Figure 4.2 Response of the Product Concentration for ± 20% steps in Feed Flow (DDC feedback control)

have one set of constants that could be used over a range of sampling times, the constants used in DDC14 were selected as the base case, Table 4.1.

Table 4.1

Feedback Controller Reference Condition

Control Loop Description	Poll Time (seconds)	Proportional Constant	Integral Constant (seconds)
First effect liquid level	64	0.75	2048.0
Separator liquid level	64	2.50	4096.0
Product concentration	64	1.50	2728.0

A summary of the results for a + 20 percent increase in feed flow is shown in Figure 4.3 where the control constants in the primary control loops were:

1. Fehr's constants (Run DDC5)
2. Fehr's constants on the two levels but higher gain and integral time on the product concentration (Run DDC7)
3. Constants used in base case (Run DDC14)

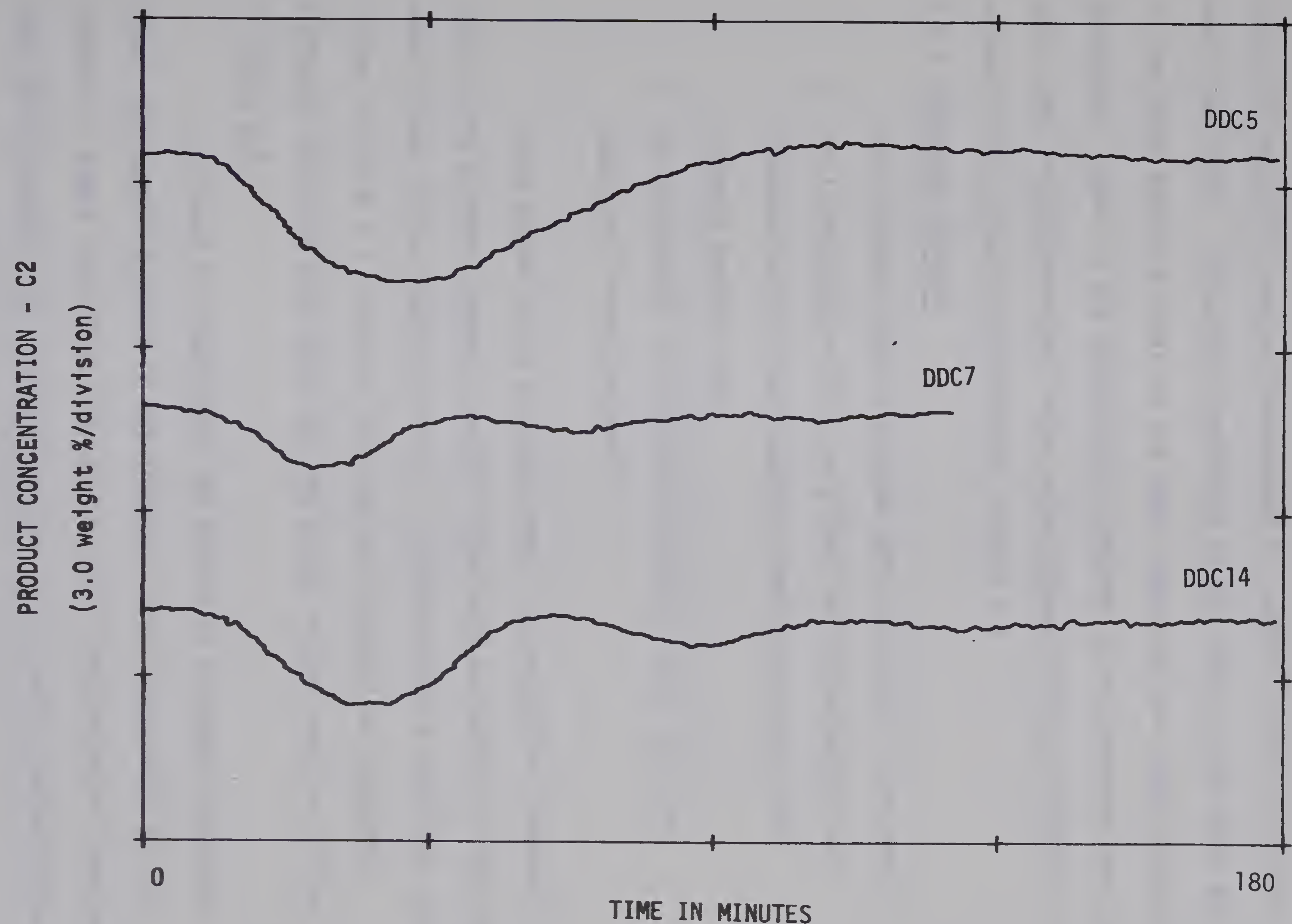
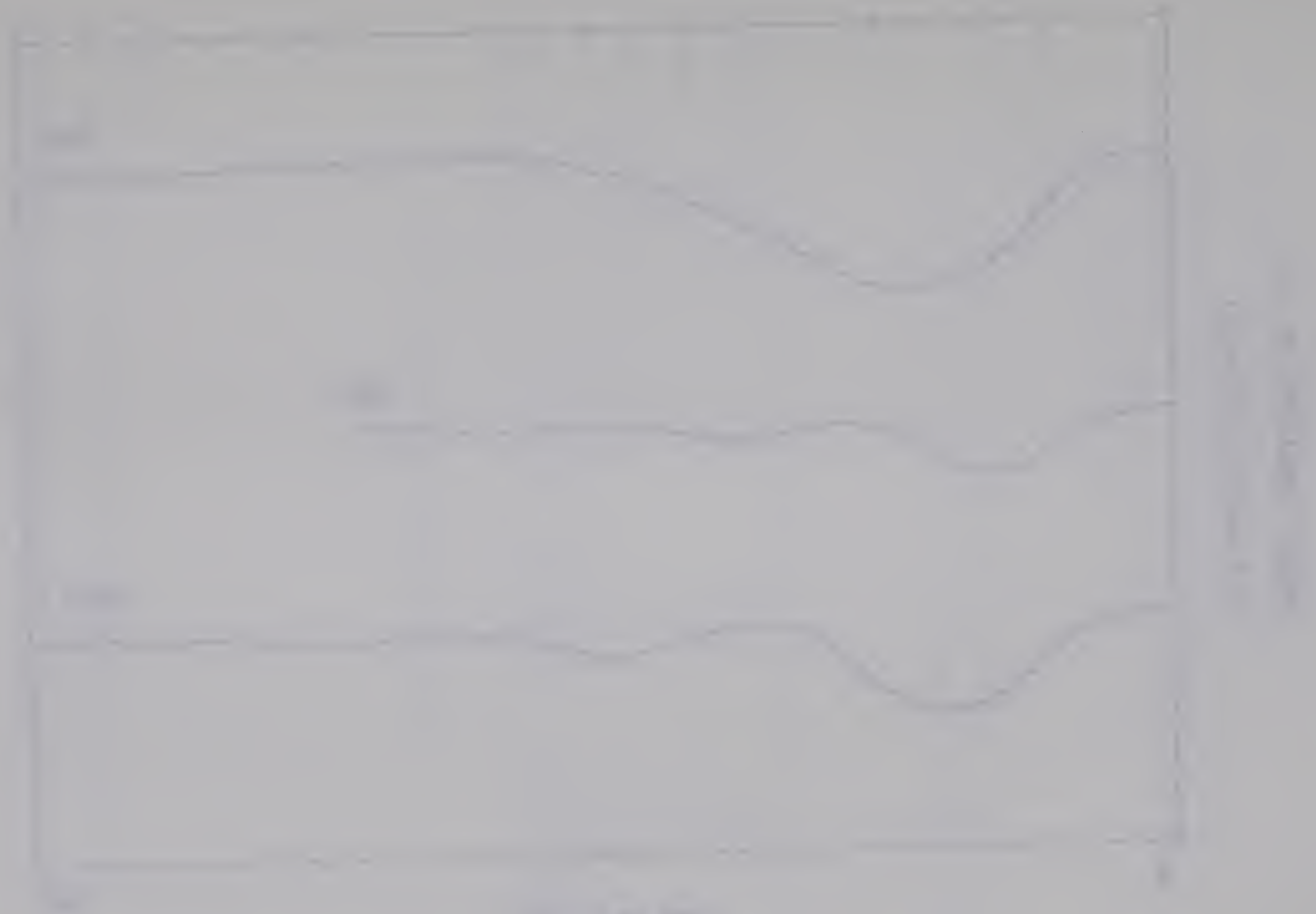


Figure 4.3 Summary of the Product Concentration Responses for ± 20% steps in Feed Flow (DDC feedback control)



Both DDC7 and DDC14 show an improvement in control relative to Fehr's base case, DDC5. The control exhibited in Run DDC7 was the best due to the lower sampling times on the level loops and the larger controller constants. Later in the experimental program it was observed that the gains used in DDC7 were too large, since steady state oscillations resulted in the product concentration, Figure 4.4.

4.3.2 Feedback Control

The feedback control studies were conducted in two parts:

1. an investigation into the effect on the evaporator performance when larger sampling times were used in the primary control loops
2. an evaluation of inferential control using the five equation linearized model.

Upon establishing the base case, the controller constants used in the primary feedback control loops were equivalent to the base case in most of the remaining runs. The "Nonreference Condition Code" column in Table 4.2 indicates the runs with controller constants different than the base case, and the actual values used are listed in Table 4.3.

In order to investigate the effect of increased sampling times, the sampling times in the primary control loops were all increased in steps from 32 to 256 seconds while the controller constants were held constant. As expected the quality of control deteriorated

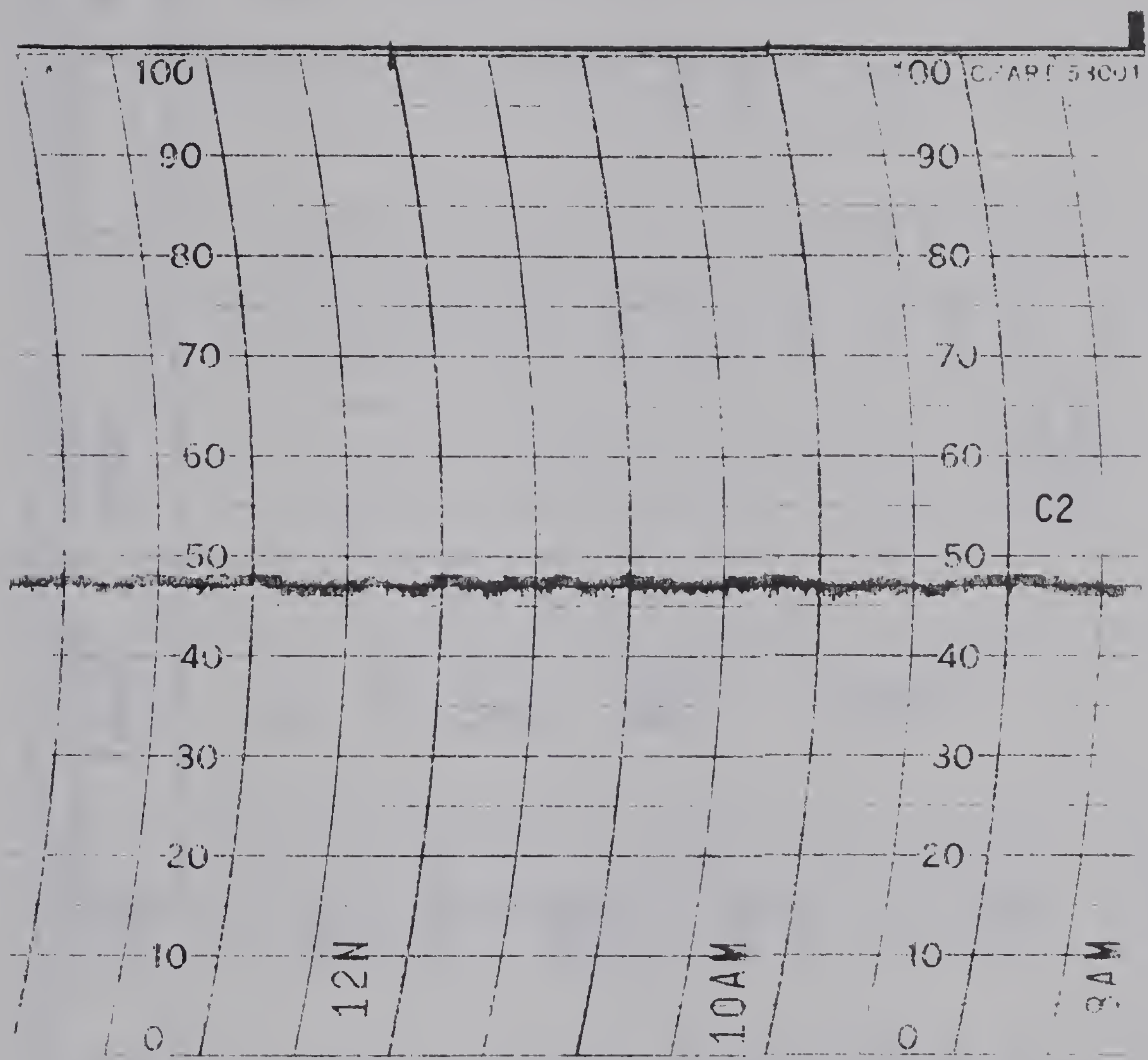


Figure 4.4 Recording of the Product Concentration Showing Steady State Oscillation Observed in Run DDC7

Table 4.2 Feedback and Feedforward Control Runs

Description of Run Mode	Run Number	Disturbance Type in Percent+ Fup	Fdown	Other	Nonreference Condition Code* ΔT	KP	KI
I P-I Feedback Control							
- Runs with Fehr's	DDC4		20		✓	✓	✓
control constants	DDC5	20			✓	✓	✓
- Tuning for Base Case	DDC1		10		✓	✓	✓
Tuning for Base Case	DDC2	10			✓	✓	✓
Tuning for Base Case	DDC3		10		✓	✓	✓
Tuning for Base Case	DDC6		20		✓	✓	✓
Tuning for Base Case	DDC7	20			✓	✓	✓
Tuning for Base Case	DDC8	20			✓	✓	✓
Tuning for Base Case	DDC9		20			✓	✓
Tuning for Base Case	DDC10		20			✓	✓
- Base Case	DDC14	20					
- Varying Sampling Time	DDC11	20			✓	✓	✓
Varying Sampling Time	DDC12		20		✓	✓	✓
Varying Sampling Time	DDC13		20		✓		
Varying Sampling Time	DDC15		20		✓		
- Inferential Control	INF1	20			✓		
Inferential Control	INF2			CF-16.7	✓		
Inferential Control	INF3			CF-16.7	✓		
II Feedforward Control							
- FF based on B1	FF1	20			✓		
- Multivariable FF using	FF2	20			✓		
square $\underline{B}$	FF5	20			✓		
- Multivariable FF using	FF3	20			✓		
least squares							

+See Table 3.3

*See Table 4.1 and 4.3 for reference and revised values

Table 4.3 Control Constants for Primary Control Loops

Run Number	First Effect Liquid Level ΔT (sec.)	K_P	τ_I (sec.)	Separator Liquid Level ΔT (sec.)	K_P	τ_I (sec.)	Product Concentration ΔT (sec.)	K_P	τ_I (sec.)
DDC4-DDC5	2	1.0	2048.0*	2	3.0	2048.0	64*	.3125	682.0
DDC1-DDC3	2	1.0	2048.0*	2	3.0	2048.0	64*	3.125	2728.0*
DDC6-DDC7	2	1.0	2048.0*	2	3.0	2048.0	64*	3.125	2728.0*
DDC8	64*	1.0	2048.0*	64*	3.0	2048.0	64*	3.125	2728.0*
DDC9	64*	1.0	4096.0	64*	3.0	2048.0	64*	3.125	2728.0*
DDC10	64*	.875	4096.0	64*	2.875	4096.0*	64*	1.25	2728.0*
DDC11	256	1.0	16400.0	256	2.5*	16400.0	256	1.25	10912.0
DDC12	256	.75*	16400.0	256	2.5*	16400.0	256	1.00	10912.0
DDC13	128	.75*	2048.0*	128	2.5*	4096.0*	128	1.5*	2728.0*
DDC15	32	.75*	2048.0*	32	2.5*	4096.0*	32	1.5*	2728.0*
INF1	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*
INF2	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*
INF3	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*
FF1	4	.75*	2048.0*	4	2.5*	4096.0*	64	1.5*	2728.0*
FF2	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*
FF5	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*
FF3	60	.75*	2048.0*	60	2.5*	4096.0*	60	1.5*	2728.0*

*Reference value

as the sampling time was increased as can be seen from the increasingly oscillatory response of C2 in Figure 4.5. This was attributed to an effective time delay which results when a continuous process is discretely sampled and is equivalent to approximately one-half the sampling time (17). This phenomenon was previously observed when the sampling times on the two level controllers were increased to 64 seconds (Runs DDC8-9 in Figure 4.2 versus DDC7 in Figure 4.3) confirming the "a priori" expectation that a control interval of one minute or larger would be marginal. Runs were also attempted at a sampling time of 256 seconds (Figures A-21 and A-22), however, control on the separator level was temporarily lost during each of the runs. The discrete nature of the response curves, in Figure A-21, for W1, B1, W2, B2 and SIFB, was caused by the 256 second sampling time. The data for the transient responses was accumulated every minute, but since the data was obtained from the actual DDC control loops the measurements of W1 and W2, and the setpoints for B1, B2 and SIFB only changed every 256 seconds resulting in the discrete responses. This was not observed on C2 since the data was accumulated from a data acquisition loop (7) having a sampling time of 64 seconds.

Inferential feedback control was implemented using the general control program which executed the control law of equation (4-2) using $\underline{\underline{KFF}} = \underline{\underline{0}}$. Although all the output variables could have been inferentially controlled, only the product concentration was controlled in this manner. The Fortran program essentially replaced the product concentration con-

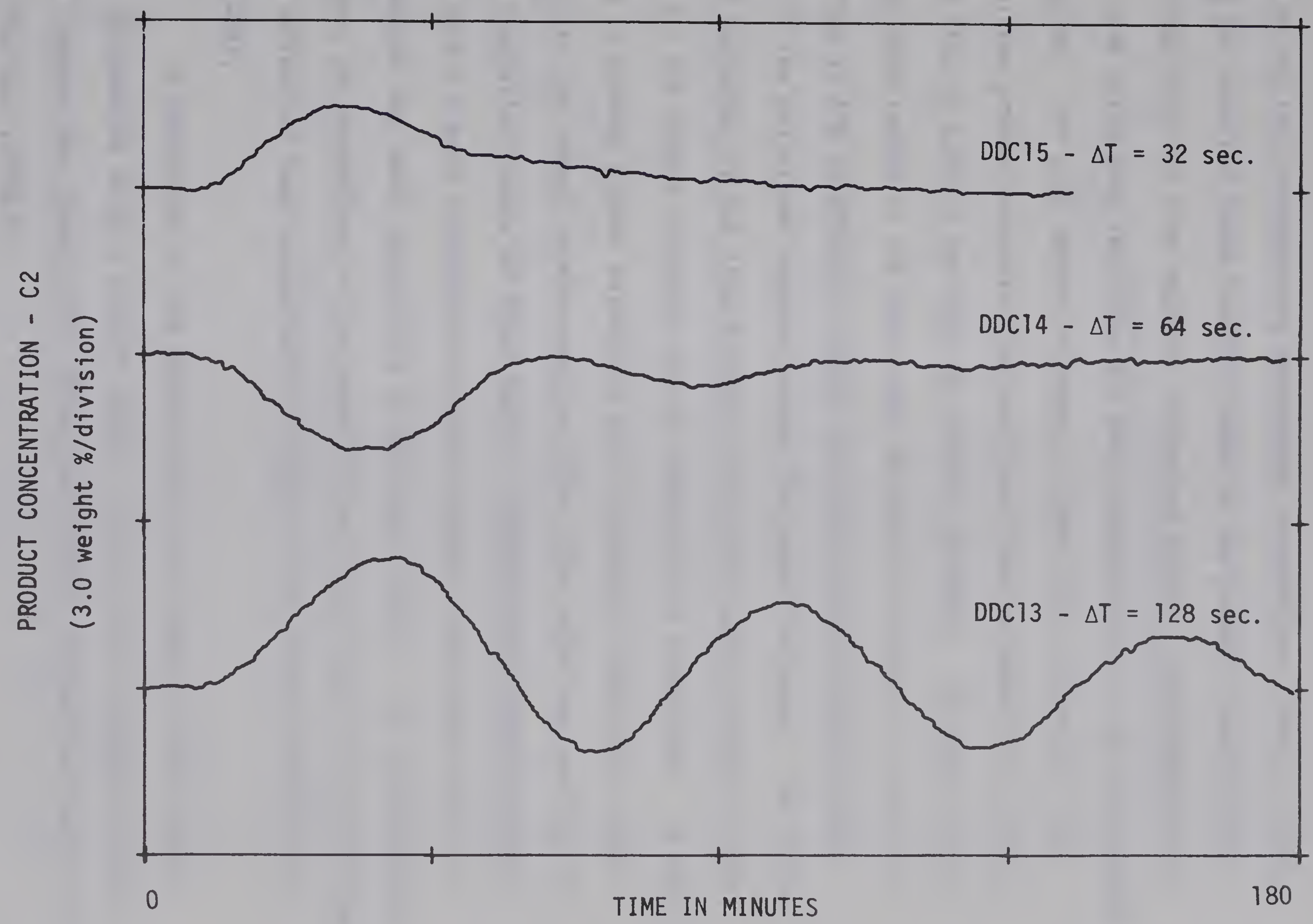


Figure 4.5 Response of the Product Concentration for $\pm$ 20% steps in Feed Flow (DDC feedback control using various sampling times on the primary control loops)

troller and the corresponding measurement transmitter, since it calculated the required steam setpoint based on the product concentration predicted from the five equation linearized model. Experimental runs were made using both feed flow and feed concentration as the disturbance variables. For a 20 percent increase in feed flow, Figure 4.6, the calculated product concentration predicted from the model (+ is model, solid line is actual) was used for control purposes. The small oscillation experienced in the base case, Run DDC14 in Figure 4.3, was not observed in the inferential control run which was attributed to the lead of the calculated response versus the actual response. The calculated responses of the liquid levels, W1 and W2, were included to show that all the output variables can be inferentially controlled. As indicated in Chapter III, the calculated level responses showed poor agreement with the actual levels when the fifth order model was used due to the integrating nature of these terms. Therefore, the responses shown in Figure 4.6 were calculated using the seventh order model which incorporates the level controllers in the state vector. For concentration changes, the deviations in the transients were small due to the relatively small effect of feed concentration changes on the process (Figures A-27 and A-28).

A comparison of the above results with those of Fehr, which were implemented using a simpler model, indicates that the more complex model reduces the steady state offset and permits inferential control of more than one variable.

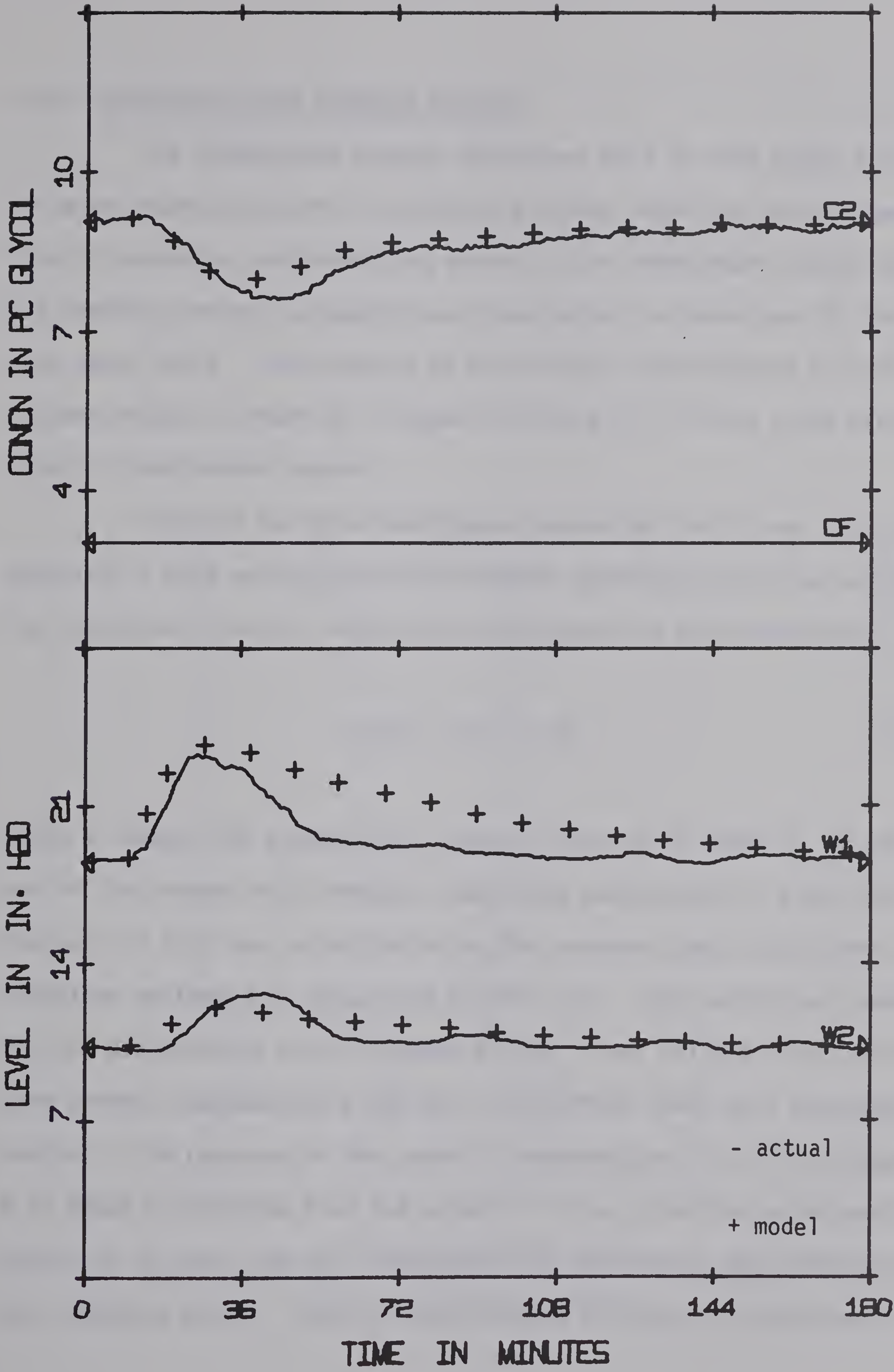


Figure 4.6 Response of the Output Variables for Run INF1 (+20% step in F under feedback control with inferential control on C2)

4.3.3 Feedforward and Feedback Control

The feedforward control techniques used in this study also contained feedback control to eliminate offset resulting from unknown load disturbances, and modelling errors in the feedforward controllers. The feedback control parameters were equivalent to those used in the base case, DDC14. The response of the product concentration to a 20 percent change in feedflow is shown in Figure 4.7 for the three different modes of feedforward control.

A single variable feedforward controller for C2 was used to establish a base so that the multivariable techniques could be evaluated. The feedforward control action was calculated from the relationship

$$\Delta SIFB = 0.61 * \Delta B1 \tag{4.8}$$

using a "dummy" DDC proportional control loop, which added to the output of the conventional feedback loop which controlled the steam flow. The gain of 0.61 was calculated using the process steady state data and relations analogous to those used by Fehr (3). This controller compensated for any disturbances which affected B1 (eg. F and TF) and also provided some dynamic compensation since the first effect level uses averaging control. The response of the product concentration, Run FF1 in Figure 4.7, shows a deviation from the setpoint in the direction which would result in an open loop run, indicating the feedforward gain should have been somewhat higher. Fehr's results using this type of feedforward

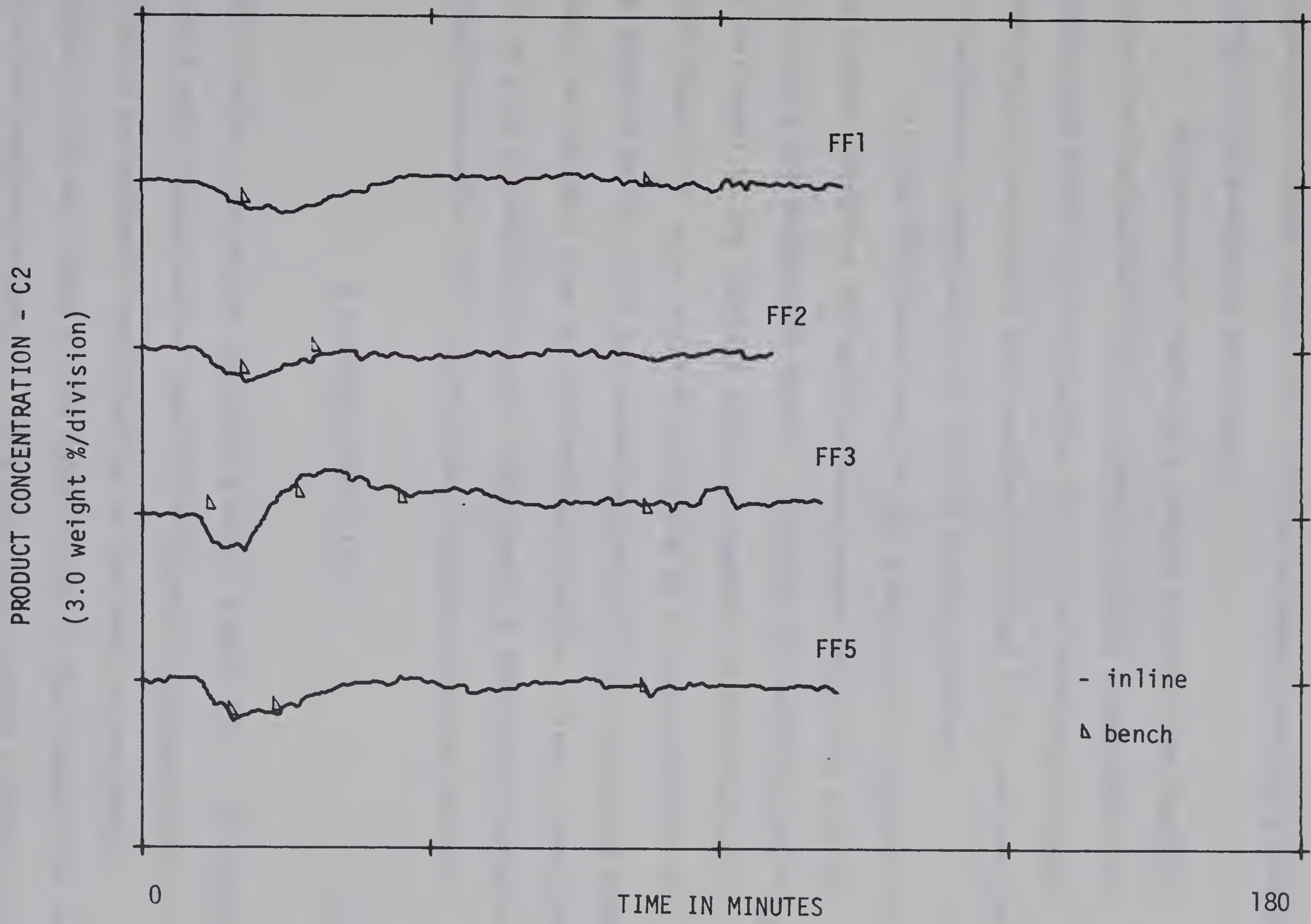
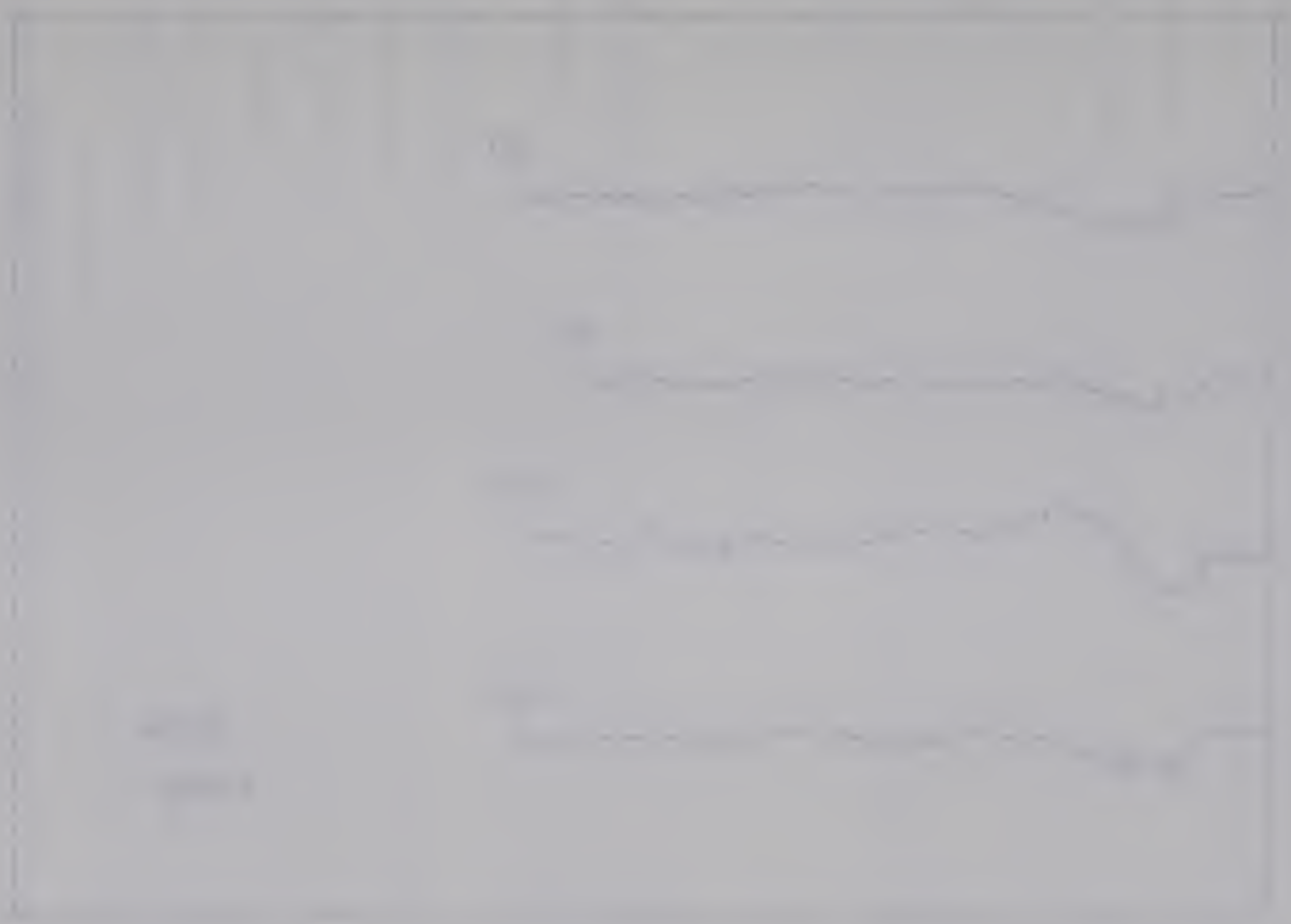


Figure 4.7 Response of the Product Concentration for + 20% steps in feed flow (Feedback and feedforward control)



compensation showed no deviation from the setpoint indicating some small change in the evaporator performance.

Multivariable feedforward control and multiloop feedback control was implemented using the general control program and the generalized control law of equation (4.2). The feedforward control matrices were calculated from equations (4.6) and (4.7), and resulted in feedforward compensation for all the output variables.

In the fifth order model of the evaporator the $\underline{\underline{H1}}$ matrix is not square, therefore the weighted least square method was used for calculating the feedforward matrix. Initially the weighting matrix $\underline{\underline{Q}}$ was set equal to the identity matrix $\underline{\underline{I}}$, however the experimental results showed this to be unsatisfactory due to a large deviation from the setpoint on C2. This was caused by the small gain resulting between steam flow and feed flow in the feedforward matrix. Since it was difficult to pick the weighting matrix intuitively, a sensitivity analysis on the state vector (18) indicated the following weighting matrix:

$$\underline{\underline{Q}} = \mathrm{diag}(10\ 100\ 1\ 10\ 1) \tag{4.9}$$

would provide a reasonable feedforward matrix (Table 4.4). The response of the product concentration, Run FF3 in Figure 4.7, as measured from the inline refractometer went opposite to the bench refractometer readings. This was caused by the sudden change in the product flow rate due to the feedforward control. This effectively caused a change in the

Table 4.4

Feedforward Control Matrices

Weighted Least Squares Technique

$$\underline{\underline{KFF}} = \begin{bmatrix} 1.248 & 0.481 & 0.0 \\ 1.721 & 1.928 & 0.0 \\ 1.586 & -2.860 & -0.4627 \end{bmatrix}$$

Inverse Technique

$$\underline{\underline{KFF}} = \begin{bmatrix} 1.0 & 0.2456 & 0.0 \\ 1.0 & 1.0 & 0.0 \\ 1.0 & -.4868 & 0.0 \end{bmatrix}$$

product temperature which was not immediately compensated for, by the in-line refractometer. The bench readings showed that the process exhibited a wrong-way response indicating the coefficient for feed flow to steam in the feedforward matrix was too high. This also affected the levels which is apparent from Figure A-31. The gain could have been reduced by changing the diagonal elements in the weighting matrix, $\underline{\underline{Q}}$. An increase in the coefficient effectively puts less weighting on the corresponding state variable resulting in a lower gain, however, the other gains in the matrix change and tend to increase.

In order to calculate a feedforward control matrix using equation (4.6), the fifth order model was reduced to a third order model assuming:

1. concentration change in the first effect is small
2. liquid enthalpy change can be neglected,

resulting in a square $\underline{\underline{H1}}$ matrix. Although these assumptions are not physically realistic, the reduced model still gives a good prediction for C2 and the response of the product concentration using this feedforward matrix (Table 4.4) was comparable to the other two techniques, Runs FF2 and FF5 in Figure 4.7. The levels only showed a small deviation in these runs (Figures A-30 and A-32).

In general, all the feedforward control systems resulted in better control since the responses on the product concentration were faster and showed a smaller deviation from the setpoint compared to the base case. Although there was no appreciable difference in the response

curves for the various feedforward control methods, the multivariable controllers perform feedforward control for a number of controlled variables and compensate for all the measurable load variables. However, a mathematical model is required and a computer (or equivalent device)is needed in order to implement the method.

4.4 Conclusions

1. Using DDC feedback control, the response of the product concentration for load disturbances showed an improvement over the previous results (3) even when a sampling time of 64 seconds was used in the primary control loops.
2. Under DDC feedback control with constant controller constants, the quality of control deteriorates as the sampling time on the primary control loops is increased. The design rule of sampling times equal one-tenth of the dominant system time constant can be used as a guide but must be adjusted to allow for interactions in multivariable systems.
3. Using the five equation linearized model for inferential feedback control on the product concentration, results in a workable control technique and showed control comparable to conventional feedback control.
4. All three feedforward control techniques significantly improved the response of the product concentration for load changes, however, no significant difference was observed for the different feedforward controllers.

CHAPTER V

MULTILOOP/MULTIVARIABLE CONTROL

5.1 Introduction

Special algorithms for multivariable feedforward, inferential, and predictive control are not yet part of commercial digital control (DDC) systems such as the one used at the University of Alberta. Therefore in order to investigate these techniques, it was necessary to develop a special control program for the University's IBM 1800 computer which enabled multivariable control techniques to be used in conjunction with or in place of DDC.

Since the control program was not core resident, it had to be transferred from bulk (disk) storage at each control interval and therefore, an excessive amount of computer execution time would result if relatively small sampling times were used. However, previous work using conventional DDC feedback control algorithms demonstrated significantly poorer control when larger sampling times were used (Chapter IV). This effect was expected since the sampling and zero hold inherent in digital control is approximately equivalent to introducing a pure time delay, of one-half the sampling interval, into the equivalent continuous system (17). Thus it was essential to develop an algorithm that would compensate for this sampling effect and permit the use of control intervals greater than one minute.

A generalized, modular, Fortran program was then designed and implemented on the IBM 1800 so that the more advanced control techniques could be developed. The program performed state estimation, model calculations, control calculations and data accumulation using the basic assumption that a state space mathematical model of the process was available. A series of experimental runs were then made to test the program and demonstrate the advantage of more advanced computer control techniques.

5.2 Theory

The use of a digital control computer, rather than conventional instrumentation, to control a continuous process results in an additional dynamic element in the control loop due to the digital interface. Moore (17) showed that the dynamic contribution of the digital interface can be approximated by a pure time delay with following magnitude:

$$TD = \frac{1}{2}\,\Delta T \tag{5.1}$$

He developed a modified proportional plus integral control algorithm which used a process model, consisting of a first order lag and a pure time delay, to compensate for the effective time delay of the digital interface and process.

This approach was extended to multivariable systems, based on a standard state space model of the process to be controlled, resulting in a proportional plus integral control algorithm which can

be expressed in matrix notation as:

$$\underline{U}(n) = \underline{\underline{KP}}(\underline{YC}(n+\tfrac{1}{2}) - \underline{R}(n)) + \sum_{i=0}^{n} \underline{\underline{KI}}(\underline{Y}(i)-\hat{\underline{YC}}(i)) \qquad (5.2a)$$

$$\underline{R}(n) = \underline{\underline{K}}\ \underline{SP}(n) \qquad (5.2b)$$

The proportional mode uses an "analytical predictor" on the output variables in an attempt to eliminate the effective dead time of the interface by analytically predicting the output variables at one-half sampling time in the future. Therefore, control action is based on this predicted value instead of on the actual value. The estimate of the output variables at one half sampling time in the future was made using the general difference equation form of the state space model of the process:

$$\underline{X}(n) = \underline{\underline{\Phi}}\ \underline{X}(n-1) + \underline{\underline{H1}}\ \underline{U}(n-1) + \underline{\underline{H2}}\ \underline{D}(n-1) \qquad (5.3a)$$

$$\underline{Y}(n) = \underline{\underline{E}}\ X(n) + \underline{\underline{F}}\ \underline{V}(n) \qquad (5.3b)$$

Since the model of the process had a time basis equal to the control interval, the output variables were predicted at time $n+\frac{1}{2}$ by linearly combining the predicted output variables at time n+1 with the best estimate at time n. The modified integral mode was designed to eliminate "reset windup" on setpoint changes. The effective setpoint for the

integral mode, $\hat{\underline{YC}}$, was calculated from the model using:

$$\hat{\underline{YC}}(n) = \underline{\underline{E}}\ \underline{\underline{\Phi}}\ \underline{X}(n-1) + \underline{\underline{E}}\ \underline{\underline{H1}}\ \underline{UP}(n-1) + \underline{\underline{F}}\ \underline{V}(n) \qquad (5.4)$$

and is based on the previous value of $\underline{X}$, the current value of $\underline{V}$ and the previous proportional control contribution. Therefore on setpoint changes the integral mode is "inactive", providing the model is accurate, and only proportional control results. In order to avoid offset, inherent when only proportional control is used, the desired setpoint of the process, $\underline{SP}$, was modified using equation (5.2b) where $\underline{\underline{K}}$ is essentially the inverse of the process closed loop gain (see Appendix B for derivation), and $\underline{R}$ is the setpoint used in the actual control law (equation 5.2a).

Process time delays are compensated for in an analogous manner by predicting the output variables over a time equivalent to the process time delay plus one-half the control interval. Since there are no significant process time delays on the evaporator, this was not implemented.

A general multivariable control law of the form:

$$\underline{U}(n) = \underline{\underline{K1}}\ \text{YS}(n) + \underline{\underline{K2}}\ \underline{XS}(n) + \underline{\underline{K3}}\ \underline{XC}(n) + \underline{\underline{K4}}\ \underline{DM}(n)$$

$$+ \underline{\underline{K5}}\ \underline{VM}(n) + \underline{\underline{K6}}\ \underline{SP}(n) + \underline{\underline{K7}}\ \underline{UI}(n-1) \qquad (5.5)$$

was then formulated (Appendix B) which incorporated the following control algorithms:

1. proportional control with prediction
2. conventional and modified integral control
3. feedforward control
4. option of using inferential control, measured variables or a linear combination of the two.

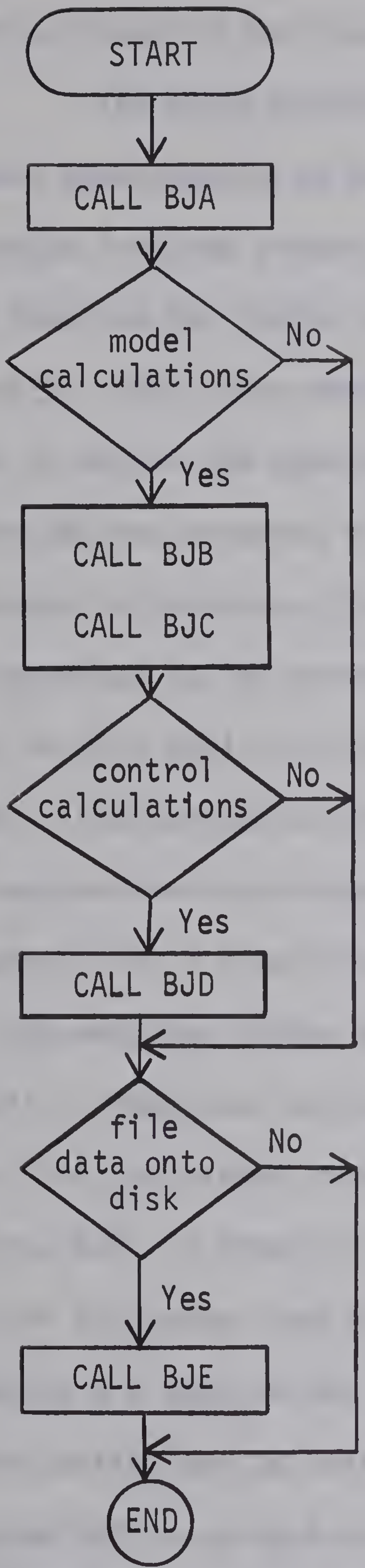
A parallel study in the department developed multivariable proportional, integral and feedforward controller matrices for use in equation (5.5) based on optimal control theory (19).

5.3 Multivariable Control Program

A general control program was written to facilitate the development and implementation of the multiloop and multivariable control methods. The program was intended as an aid for research work within the department rather than as a commercial software system. The total system was written in Fortran IV and was implemented on the IBM 1800 TSX System. The program is general so that it can be used on any process providing a mathematical model (of the form given in equation 5.3) is available and of course a compatible computer system.

A simplified flow diagram of the control program is given in Figure 5.1. The basic system is comprised of a number of modules which perform specific functions. The combination of these modules results in the three main applications of the program:

1. data accumulation
2. open loop model calculations
3. control calculations and output



Comments:
Start on timer interrupt and obtain data from disk files

STATE ESTIMATION
- Obtain current estimates of $\underline{X}$, $\underline{U}$, $\underline{D}$, $\underline{V}$, $\underline{Y}$

MODEL CALCULATIONS
- Convert variables to model units
- Perform model calculations to generate $\underline{Y}(n)$, $\underline{Y}(n+1/2)$ etc.

CONTROL CALCULATIONS
- Control based on either DDC algorithms or algorithms within the control program (eg. combined algorithm)
- Send output to measurement or setpoint of DDC loop

DATA FILING
- File calculated and/or measured data into a disk file

File information and parameters required for next iteration and exit

Figure 5.1 Simplified Flow Diagram of Control Program

which as shown in the flow diagram can be used independently or combined.

The state estimation section, BJA, is used to obtain the current measurements of the state, manipulated, disturbance and output variables from the process being controlled. The measurements can be read from the DDC tables or can be obtained by reading the multiplexers directly. The latter method requires a number of conversion steps in order to obtain the measurement in engineering units. The model calculation section consists of two modules; measurement conversion, BJB, and model calculations, BJC. The measurements used in the model equations are converted to the corresponding model form (normalized perturbation units in this application) before the state and output vectors are calculated. Incorporated at this point is the ability to average or filter the measured and calculated state and output variables. The form used (Chapter IV) is a simplified form of Kalman filtering and is used to estimate non-measured states and also used in the implementation of inferential control. Prediction of the output variables in the future is also available. The calculated state and output vectors are used in the control section, BJD, in order to implement the various control algorithms. Control can originate from the actual DDC loops, in which case, the output variables are sent to the measurements of the corresponding DDC loop, or control action can be calculated within the program. The control action is converted to an equivalent value in engineering units before it is sent to the setpoint of the manipulated variables' control loops. The final section is the data filing module, BJE, which files measured and/or

calculated variables into a disk file for access at a later time.

A number of design considerations had to be dealt with when writing the control program. Due to the structure of the IBM 1800 (3,19,20), a large number of different applications exist and only programs of major importance can be permanently stored in core. Therefore the control program was stored on disk in an executable core image format. Once the program was initialized, it was automatically transferred to core every control interval and executed. An updating of measurements, control action and data files was performed at each control interval and continued until the software timer, used to repetitively queue the program, was turned off. The restricted size of the core available for program execution (9.7K) necessitated using exact dimensions on the vectors and matrices, and combining several calculation steps into one. This conserved core space and ultimately lowered the execution time since the number of core overlays (which requires read-writes to disk) were minimized. Disk read-writes were minimized particularly in the data filing module where the data, at a particular control interval, were filed such that only one write statement was required. The data were sorted in an offline program in order to obtain the hard copy documentation, punched data cards and graphs. All input data checks and calculations, required to initialize the control program, were handled by another program which could be run on a lower priority than the actual control program. This enabled unrestricted execution of the control program. The lower priority program was also used for

communication between the operator, computer and process.

The more detailed program documentation and user's manual is available in reference (24).

5.4 Discussion of Results

A series of experimental runs were conducted in order to evaluate the control techniques presented in this chapter and to test the general control program. The runs have been summarized in Table 5.1 and the control constantsused in the primary control loops are in Table 5.2.

The use of prediction on the output variables was evaluated by using the normal proportional plus integral control algorithms, with one-half interval prediction, and comparing the results with standard DDC runs. The control constants on the primary control loops were held constant, and as the sampling time was increased in steps from one to eight minutes, the quality of control did not deteriorate significantly, Figure 5.2. There was only a small oscillation in the product concentration response, RUN PRED4 in Figure 5.2, even at a sampling time of 8.0 minutes. Similar results were observed on the level responses which are shown in Figures A-33 to A-36. The discrete nature of the response curves is a direct result of the sampling times used, and is particularly noticeable at the eight minute sampling time.

Comparing the results with those obtained using standard DDC feedback control, Figure 5.3, shows a significant improvement in control when prediction is used. With conventional control there was a slight deterioration in control in going from a sampling time of 32 seconds to

Table 5.1

Multiloop/Multivariable Experimental Runs

	Description of Control Modes	Run Number	Disturbance Type in Percent+			Non Reference Condition Code*		
					Other	ΔT	KP	KI
I	P-I Control with Prediction	PRED1	20			✓		
	P-I Control with Prediction	PRED2		20		✓		
	P-I Control with Prediction	PRED3	20			✓		
	P-I Control with Prediction	PRED4		20		✓		
	P-I Control with Prediction	PRED5			C2SP9.3-10.2-9.3	✓		
II	Modified P-I Control	LSU1			C2SP9.3-10.2-9.3	✓		
	Modified P-I Control	LSU2	20			✓		✓
III	Combined Algorithm	COMB1	20			✓		
	Combined Algorithm	COMB2		20		✓		
	Combined Algorithm	COMB3	20			✓		

+See Table 3.3 for definition

*See Table 4.1 for reference constants

Table 5.2

Control Constants for Primary Control Loops

Run Number	First Effect Level			Second Effect Level			Product Concentration		
	ΔT (min.)	KP	τ_I (sec.)	ΔT (min.)	KP	τ_I (sec.)	ΔT (min.)	KP	τ_I (sec.)
PRED1	1.0	.75*	2048*	1.0	2.5*	4096*	1.0	1.5*	2728*
PRED2	2.0	.75*	2048*	2.0	2.5*	4096*	2.0	1.5*	2728*
PRED3	4.0	.75*	2048*	4.0	2.5*	4096*	4.0	1.5*	2728*
PRED4	8.0	.75*	2048*	8.0	2.5*	4096*	8.0	1.5*	2728*
PRED5	1.0	.75*	2048*	1.0	2.5*	4096*	1.0	1.5*	2728*
LSU1	1.0	.75*	2048*	1.0	2.5*	4096*	1.0	1.5*	2728*
LSU2	1.0	.75*	512	1.0	2.5*	512	1.0	1.5*	341
COMB1	4.0	.75*	2048*	4.0	2.5*	4096*	4.0	1.5*	2728*
COMB2	4.0	.75*	2048*	4.0	2.5*	4096*	4.0	1.5*	2728*
COMB3	4.0	.75*	2048*	4.0	2.5*	4096*	4.0	1.5*	2728*

*reference value

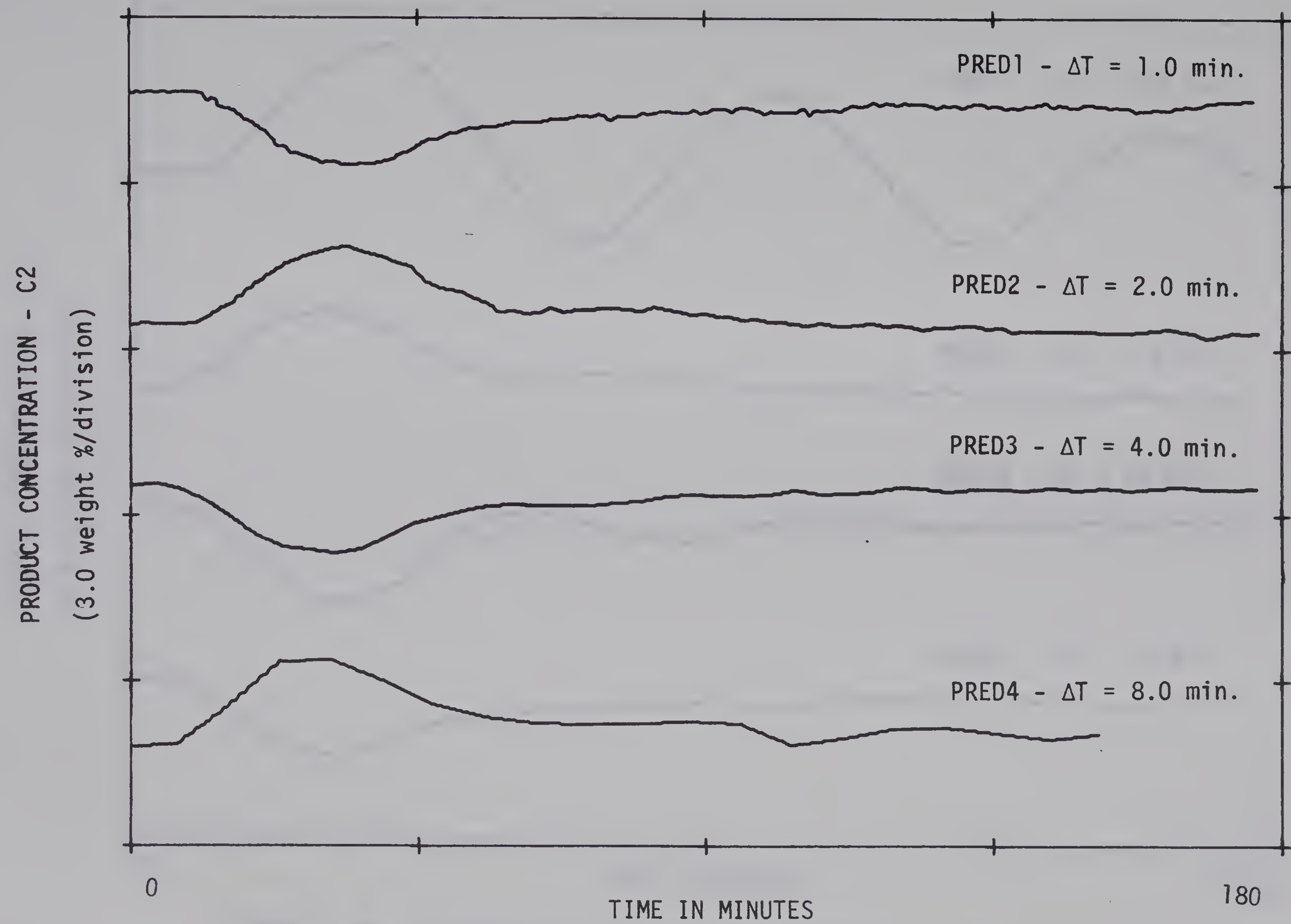
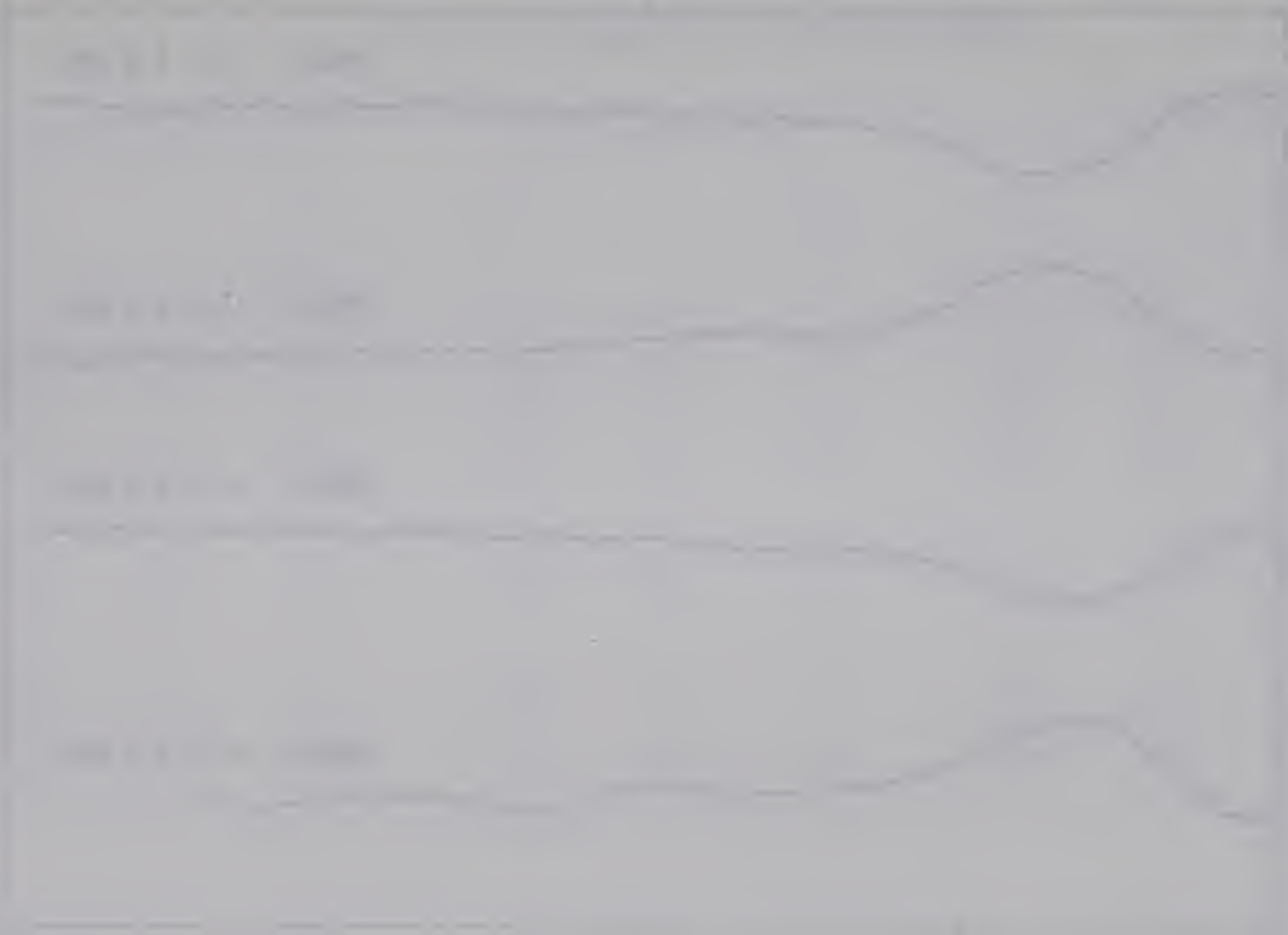


Figure 5.2 Response of the Product Concentration for ± 20% steps in Feed Flow (feedback control with prediction using various sampling times on the primary loops)



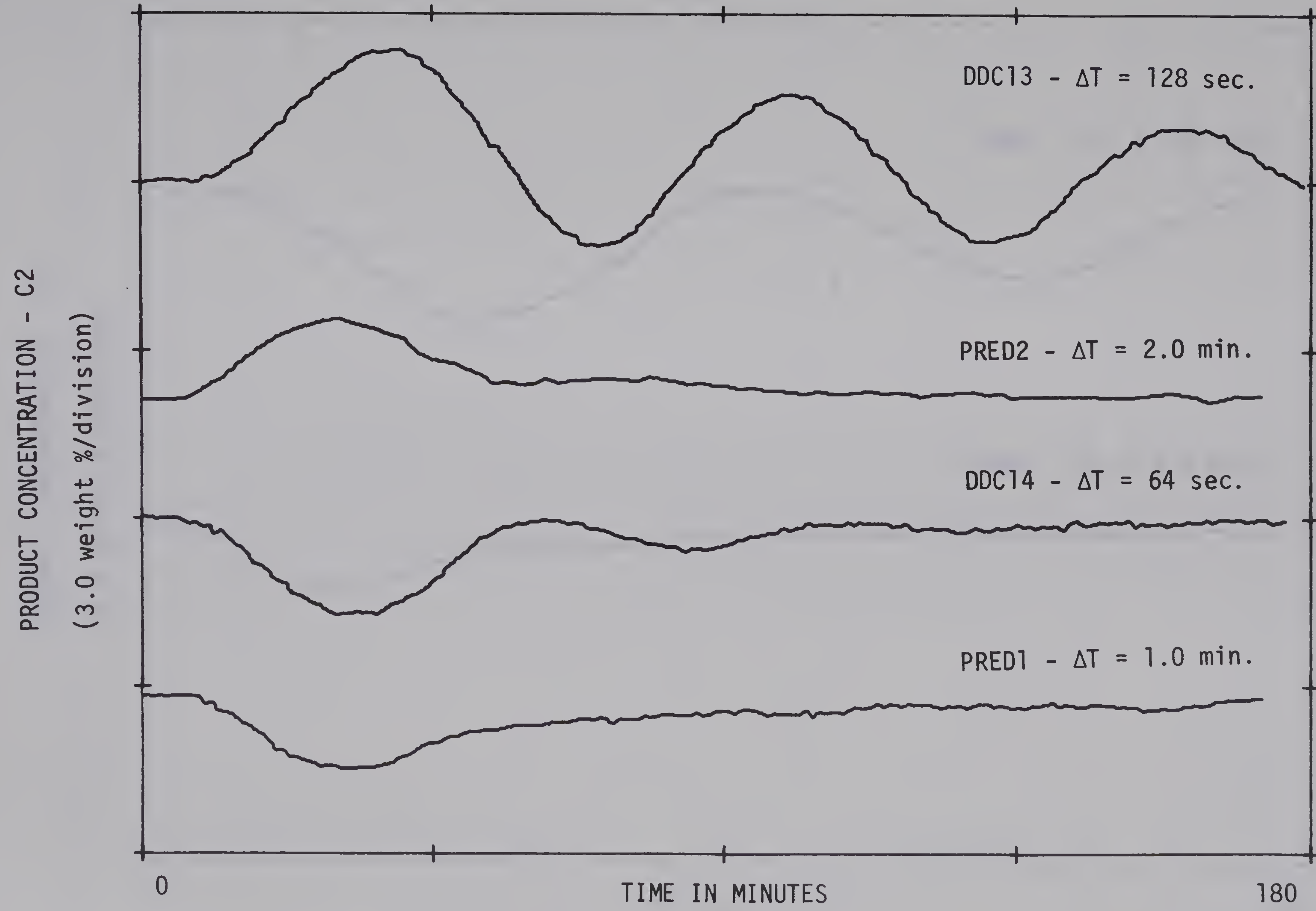


Figure 5.3a Comparison of the Product Concentration Responses for $\pm$ 20% steps in feed flow using DDC feedback control and feedback control with prediction

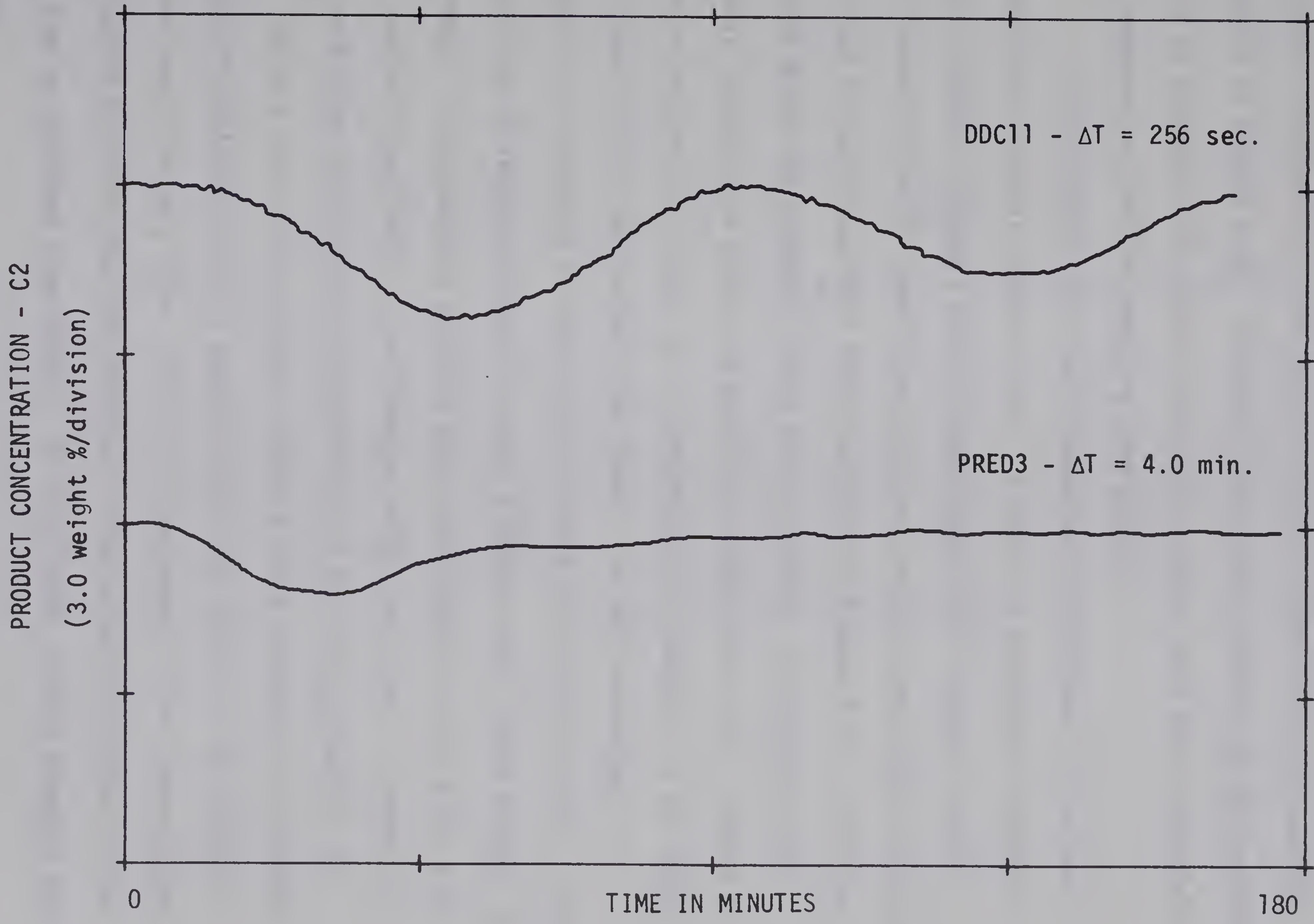


Figure 5.3b Comparison of the Product Concentration Responses for + 20% steps in feed flow using DDC feedback control and feedback control with prediction

64 seconds (Chapter IV), and control was not practical at 256 seconds, Run DDC11 in Figure 5.3b. In fact, satisfactory control of the evaporator cannot be obtained with control intervals greater than two minutes without compensating for the sampling time effect.

Both setpoint and load changes were introduced to evaluate the modified integral control mode. In order to establish a comparison for setpoint changes, normal proportional plus integral control with one-half interval prediction was used and the same control constants were used in each case, RUNS PRED5 and LSU1 in Figure 5.4. The arrows indicate where the setpoint change was initiated. The results were somewhat inconclusive since the modified integral run, LSU1, showed a faster rise time for a step up in concentration, however, it was somewhat slower for the equivalent step down. The main advantage, although not illustrated in Figure 5.4, is that no (or very little) integral action acts on setpoint changes so that problems with "reset windup" do not result. Considerable problems were encountered in tuning the modified control algorithm for load changes using feed flow. A number of runs were made using various combinations of KP and KI, however, in each case the controlled variable response was slow and/or it overshot the desired setpoint, and it appeared that the return to the desired setpoint would be very slow. This is illustrated in the Foxboro charts in Figure (5.5), which show the response of the output variables for Run LSU2 over an extended time period. At this stage, tuning attempts were stopped.

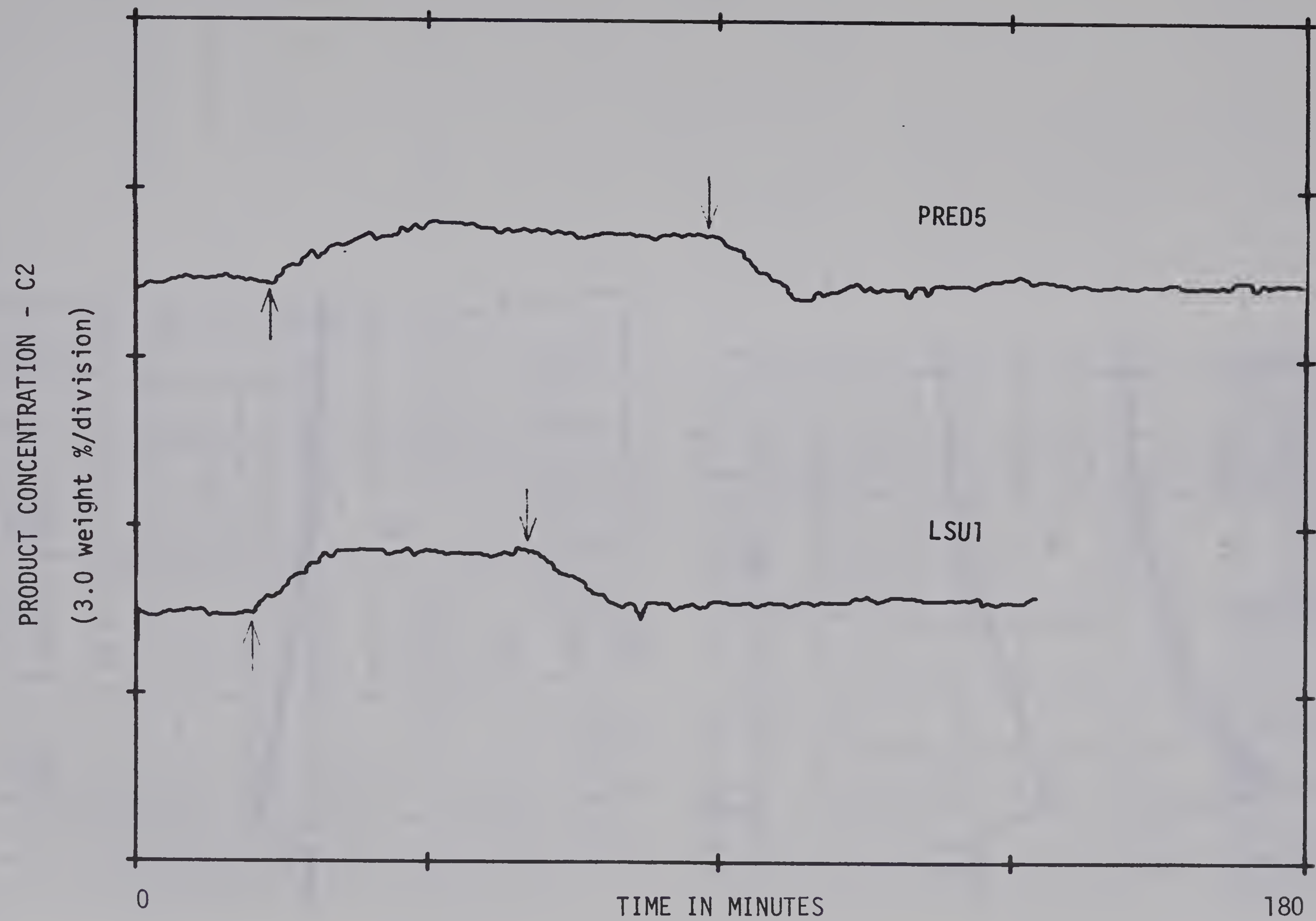


Figure 5.4 Comparison of the Product Concentration Responses for a setpoint change of 0.9 weight % using feedback control with prediction and modified feedback control

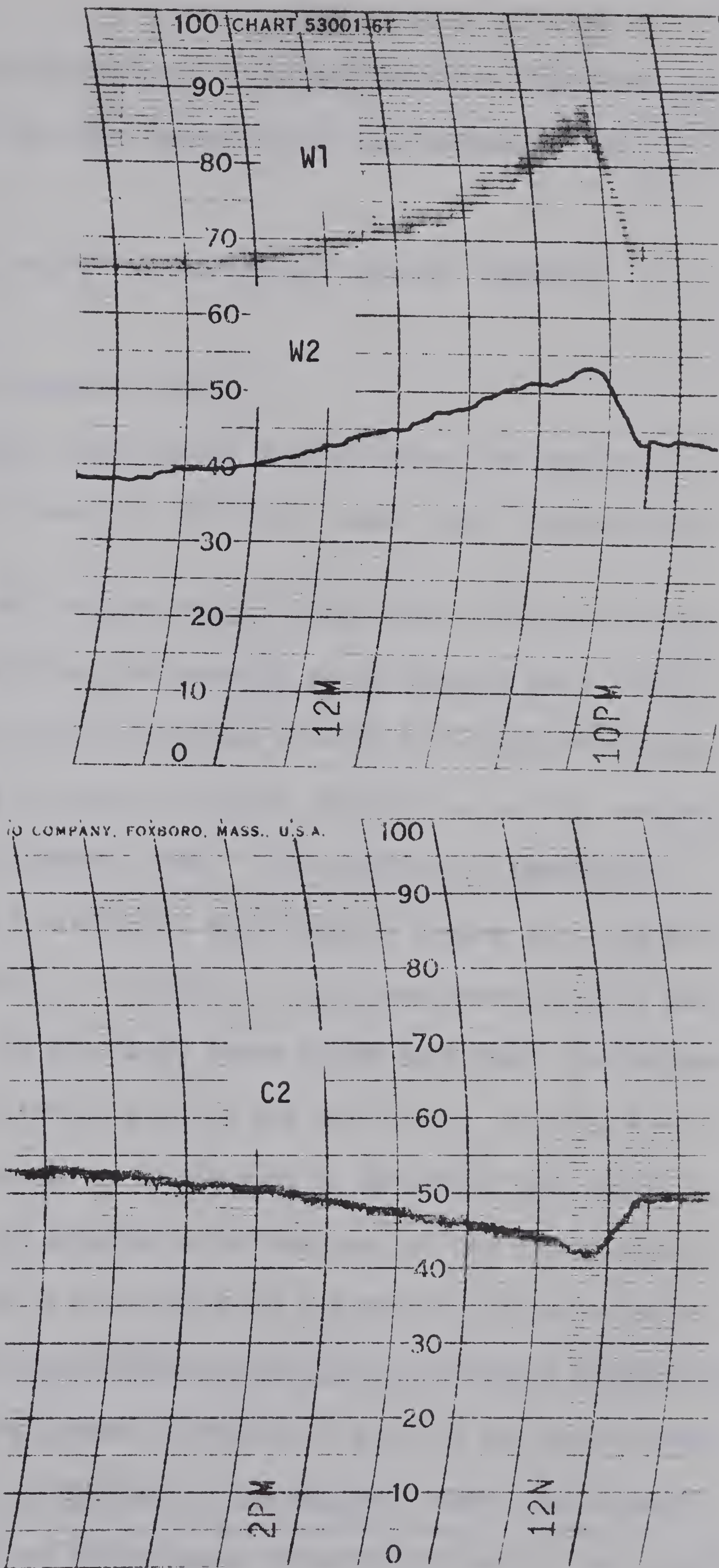


Figure 5.5 Response of the First Effect Level, Second Effect and Product Concentration for ± 20% step in feed flow (Modified P-I control mode)

The combined multiloop/multivariable control algorithm, defined by equation 5.5, was then tested using the following control algorithm options:

1. proportional control with one-half control interval prediction
2. conventional integral control
3. with and without feedforward control using the feedforward control matrix based on the third order model (Chapter IV).

The sampling times on the primary control loops were arbitrarily set at four minutes and feed flow disturbances of 20 percent were introduced, Figure 5.6. The results obtained without any feedforward compensation (COMB1) were equivalent to those obtained using DDC feedback control at a one minute sampling time. The addition of feedforward control on all the output variables, Run COMB3 in Figure 5.6, significantly improved the control. The gains in the feedforward matrix did not exactly compensate for the load, since there were small deviations in the product concentration as well as the two levels, Figures A-41 and A-42. However, these can be eliminated by adjusting the coefficients in the matrix. The apparent slow response of the manipulated and disturbance variables is indicative of the sampling time and not the control action. Also, the effect of using multivariable feedforward control can be seen by the sudden increases in each of the manipulated variables near the start of the runs. The apparent dead time between the response of the feed and manipulated variables is due to the sampling time (see Figures A-41 to A-42).

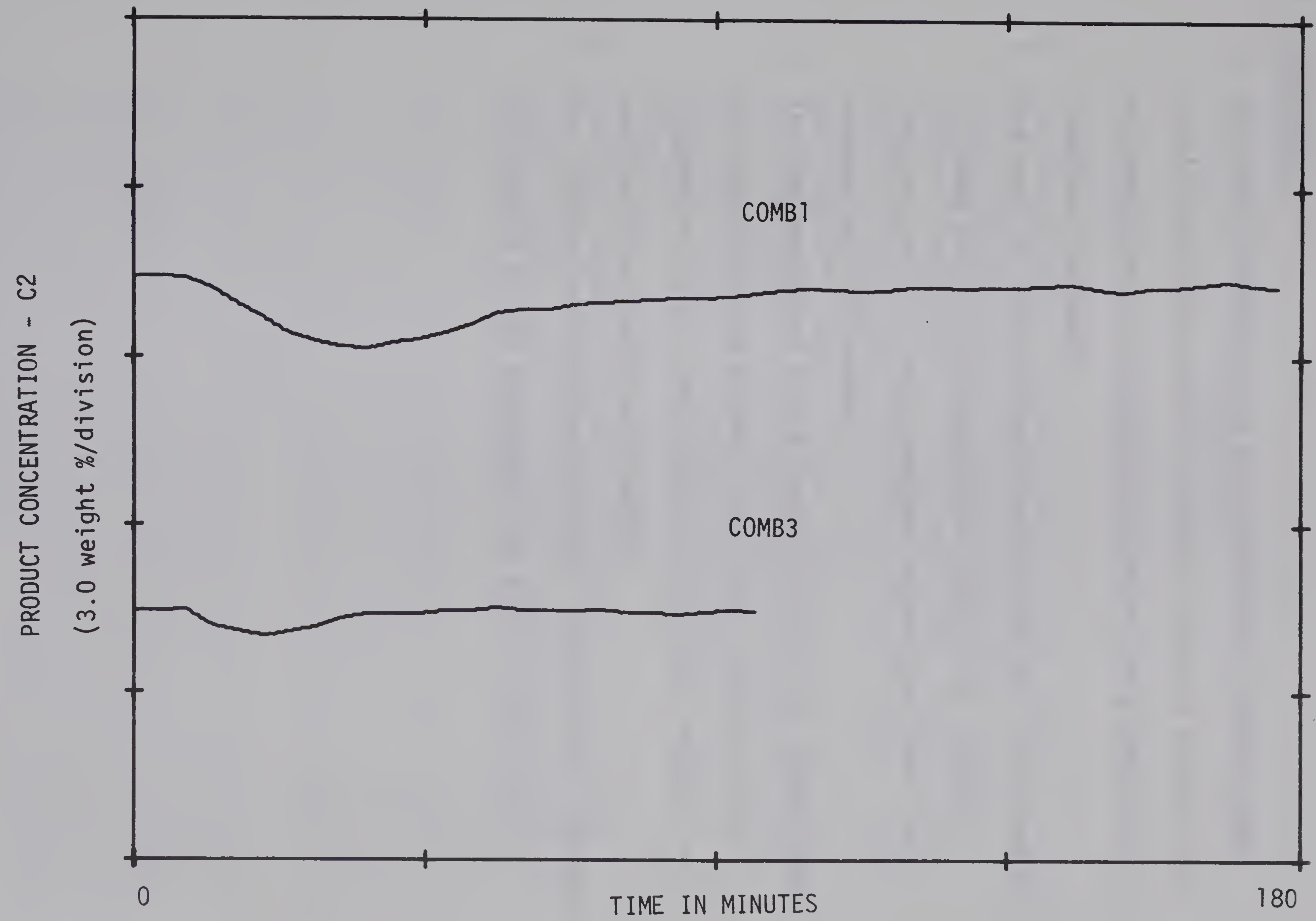
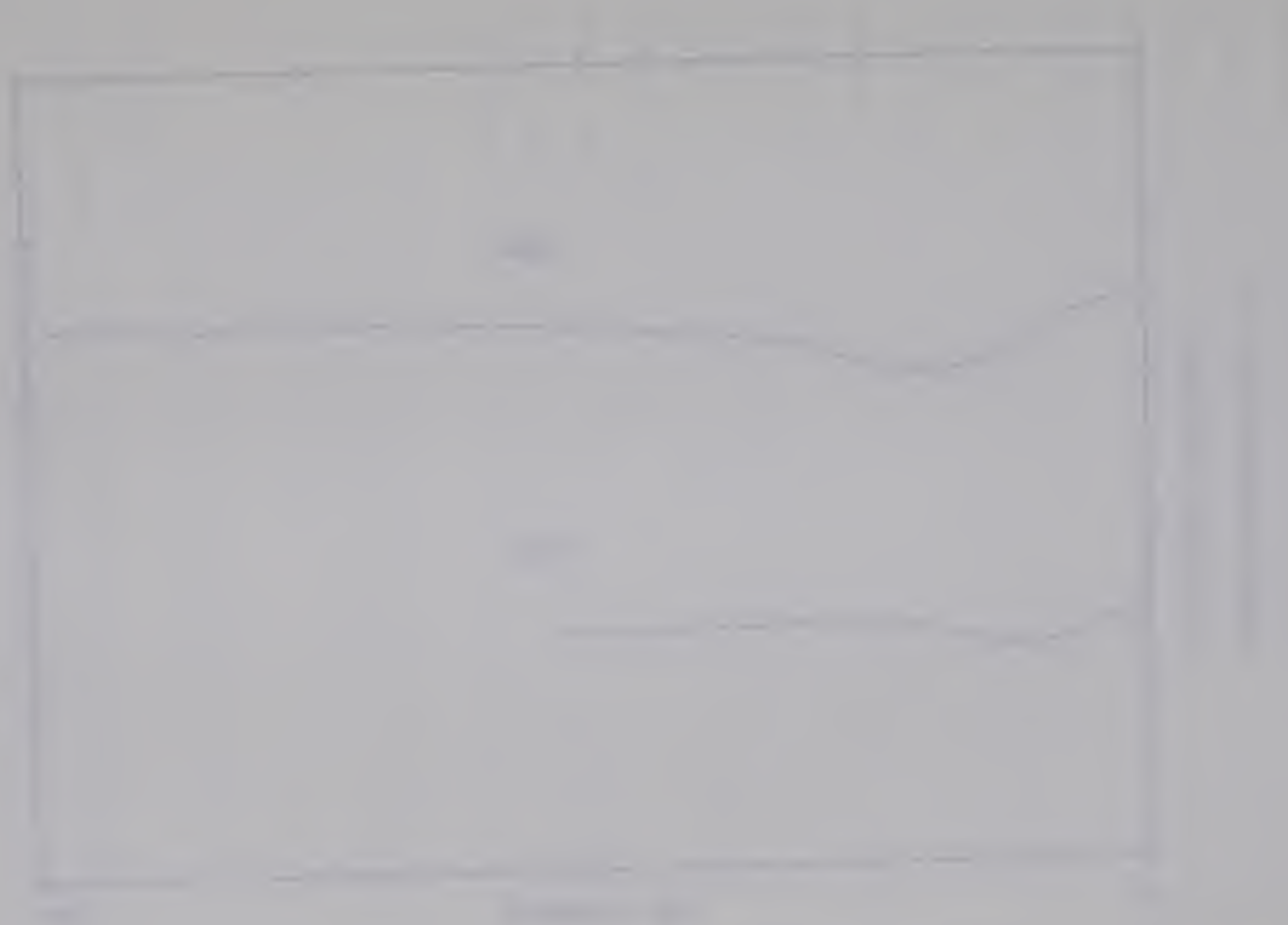


Figure 5.6 Response of the Product Concentration for +20% steps in feed flow using feedforward and/or feedback control



5.5 Conclusions

1. The use of prediction in the proportional control mode significantly improved the control, compared to runs with no prediction, as the sampling time on the primary control loops was increased.
2. The merit of the modified integral mode was left somewhat in doubt due to the inconclusive results for setpoint changes, and the tuning problems encountered when load changes were introduced.
3. The implementation of the control program was demonstrated by controlling the primary controlled variables on the evaporator using a combination of multiloop feedback and multivariable feedforward control.
4. The addition of feedforward control significantly improved the control of the evaporator for steps in feed flow.

CHAPTER VI

CONCLUSIONS

1. The five equation linearized model of the double effect evaporator satisfactorily predicted the open loop response of the product concentration for 20 percent load and manipulated variable changes about the point of linearization. Due to the integrating nature of the levels in the fifth order model, the actual level controllers had to be included in the model so that a reasonable representation could be obtained.
2. Using DDC feedback control, an increase in the primary control loops' sampling times to two minutes or higher resulted in a rapid deterioration of the overall evaporator performance. The use of the proportional control mode with prediction significantly improved the control on the evaporator and resulted in satisfactory control at a sampling time as high as eight minutes.
3. Inferential feedback control of the product concentration was implemented in a multiloop form using the five equation linearized model. The results showed the system to be workable with the quality of control being equivalent to that obtained under conventional DDC control.
4. Multivariable feedforward control of the primary controlled

variables was implemented and combined with multiloop feedback control. Two means of obtaining the steady state feedforward control matrix from the mathematical model were derived and implemented. Comparison with the more conventional single variable feed forward control, using feed flow disturbances, showed that within the limits of experimental reproducibility and noise that the three methods were equivalent.

5. The use of the modified integral mode for setpoint changes showed some promise, however, enough experimental runs were not made to draw any firm conclusion. Tuning problems were encountered when this control mode was used on load disturbances.
6. The multiloop/multivariable control program was successfully implemented on the IBM 1800 and was used to replace the DDC controllers on the primary controlled variables. All the multiloop feedback and multivariable feedforward control techniques used in the thesis were incorporated within this program, and the experimental runs which were made used the control program, demonstrating that the program was workable.
7. Although not part of the research project, operating procedures for the evaporator were established which enabled operators, who were relatively unfamiliar with the evaporator, to operate it. Also the evaporator operation was converted from a batch type process to one which was run continuously, 24 hours each day.

NOMENCLATURE

$\underline{\underline{A}}$	- coefficient matrix for linearized differential equation model
$\underline{\underline{AI}}$	- coefficient matrix for integral control mode in combined algorithm
$\underline{\underline{AX}}$	- coefficient matrix for combining calculated and measured state vector
$\underline{\underline{AY}}$	- coefficient matrix for combining calculated and measured output vector
$\underline{\underline{A1}}$	- modified $\underline{\underline{A}}$ for seventh order model
$\underline{\underline{B}}$	- coefficient matrix for linearized differential equation model
B	- bias in DDC control loop
$\underline{\underline{B1}}$	- modified $\underline{\underline{B}}$ for fifth order model
B1	- first effect bottoms flow rate
$\underline{\underline{B2}}$	- modified $\underline{\underline{B}}$ for seventh order model
B2	- product flow rate
$\underline{\underline{C}}$	- coefficient matrix for linearized differential equation model
CF	- total feed concentration
$\underline{\underline{C1}}$	- modified $\underline{\underline{C}}$ for seventh order model
C1	- first effect bottoms concentration

C2 - product concentration

C2SP - product concentration setpoint

$\underline{D}$ - model disturbance vector

DDC - direct digital control

$\underline{DM}$ - measured $\underline{D}$ vector

$\underline{DSS}$ - steady state $\underline{D}$ vector

d-p - differential pressure

$\underline{\underline{E}}$ - coefficient matrix for linearized mathematical model

$\underline{E}$ - error vector for integral control mode in combined algorithm

$\underline{\underline{E1}}$ - modified $\underline{\underline{E}}$ for fifth order model

$\underline{\underline{E2}}$ - modified $\underline{\underline{E}}$ for seventh order model

$\underline{\underline{F}}$ - coefficient matrix for linearized mathematical model

F - total feed flow rate

FF - feedforward

hF - total feed enthalpy

h1 - first effect bottoms enthalpy

$\underline{\underline{H1}}$ - coefficient matrix for difference equation model

$\underline{\underline{H2}}$ - coefficient matrix for difference equation model

$\underline{\underline{I}}$ - identity matrix

$\underline{\underline{K}}$ - coefficient matrix for modified proportional control setpoint

$\underline{\underline{KD}}$ - derivative control matrix

$\underline{\underline{KFF}}$ - feedforward control matrix

$\underline{\underline{KI}}$ - integral control matrix

$\underline{\underline{KP}}$ - proportional control matrix

$\underline{\underline{K1}}$ - matrix used in the combined control algorithm (also includes $\underline{\underline{K2}}$, $\underline{\underline{K3}}$, $\underline{\underline{K4}}$, $\underline{\underline{K5}}$, $\underline{\underline{K6}}$, $\underline{\underline{K7}}$)

M - measurement in DDC control loop

n - n^{th} time interval

O - output in DDC control loop

P-I - proportional plus integral control

PID - proportional, integral plus derivative control

Poll Time - sampling time

$\underline{\underline{Q}}$ - square weighting matrix

$\underline{R}$ - setpoint for modified proportional control mode

S - setpoint

SIFB - steam flow rate

$\underline{SP}$ - setpoint vector for output variables

TD - time delay

TF - total feed temperature (measured form of hF)

T1 - first effect bottoms temperature (measured form of h1)

$\underline{U}$ - model manipulated vector

$\underline{UFF}$ - feedforward contribution to $\underline{U}$

$\underline{UI}$ - integral contribution to $\underline{U}$

$\underline{UM}$ - measured $\underline{U}$ vector

$\underline{UP}$ - proportional contribution to $\underline{U}$

$\underline{USS}$ - steady state $\underline{U}$ vector

$\underline{V}$ - model vector

$\underline{VM}$ - measured $\underline{V}$ vector

W1 - first effect holdup

W2 - second effect holdup

$\underline{X}$ - model state vector

$\underline{XC}$ - state vector calculated from the model

$\underline{XM}$ - measured state vector

$\underline{XS}$ - combined state vector using measured and calculated state

$\underline{XSS}$ - steady-state state vector

X6, X7 - elements of the state vector in the seventh order model

$\underline{Y}$ - model output vector

$\underline{YC}$ - output vector calculated from the model

$\hat{\underline{YC}}$ - calculated setpoint for modified integral control mode

$\underline{YM}$ - measured output vector

$\underline{YS}$ - combined output vector using measured and calculated output vectors

$\underline{YSS}$ - steady state output vector

$\underline{Z}$ - "dummy" vector representing the $\int \underline{Y}\, dt$

ΔT - sampling time

$\underline{\underline{\Phi}}$ - fundamental matrix used in state difference equation

τ_I - integral time

$\underline{\ }$ - denotes a vector

$\underline{\underline{\ }}$ - denotes a matrix

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APPENDIX A

APPENDIX A

Designation of Variables

The process initial and final steady state data tables use the "process variable names" to identify each value in the table. The transient responses for any particular run define each of the response curves using the "model variable names". The correspondence between the "process and model nomenclature" is given in Table A-1 and is shown on the Evaporator schematic drawing, Figure 2.1.

Plots displaying both the actual process transients and those calculated from the mathematical model use the following notation:

- \- actual process response
- \+ model response

Experimental Run Designation

Each of the experimental runs were identified according to the type of control used on the primary controlled variables (W1, W2, C2):

OL - DDC feedback[1] control with no control on the product concentration, C2.

DDC - DDC feedback control.

INF - Feedback control using the general control program[2] with inferential control on the product concentration, C2.

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[1] Feedback control implies proportional plus integral control modes.

[2] General description of control program available in Section 5.3 or Reference (24).

FF - Feedback plus feed forward control using the general control program (exception is FF1 which uses DDC feedback control, and DDC feed forward control on C2 by manipulating SIFB based on B1).

PRED - Feedback control with prediction using the general control program.

LSU - Modified proportional plus integral feedback control algorithm using the general control program.

COMB - Feed forward and/or feedback control using the general control program.

Note: A blank column in the steady state tables indicates the process did not attain a final steady state.

Table A-1

Process Reference Steady State

Process Variable Name	Model Designation	Reference Value	Variable Description
F1	SIFB	2.00 lbs./min.	Steam flow
F2	B1	3.32 lbs./min.	First effect bottom flow
F5		1.68 lbs./min.	First effect overhead flow
F6	B2	1.64 lbs./min.	Product Flow
F7		1.69 lbs./min.	Second effect (separator) overhead flow
F8	F	5.00 lbs./min.	Total feed flow
F9		40.00 lbs./min.	Cooling water flow
F10		190.00 lbs./min.	Circulation rate
C1	CF	0.030 wt. fraction	Feed concentration
C6	C2	0.092 wt. fraction	Product concentration
L14	W1	22.0/30.0 in./lbs.	First effect liquid level
L11	W2	11.0/35.0 in./lbs.	Second effect (separator) liquid level
P20		5.5 psig	First effect pressure
P22		-15.0 in.Hg	Second effect condenser pressure
T1		98.0 deg. F	Cooling water outlet temperature
T2		222.9 deg. F	Vapor temperature in first effect
T4		182.0 deg. F	Solution temperature to second effect
T5		248.2 deg. F	Temperature of first effect steam condensate
T7	hF(TF)	190.0 deg. F	Total feed temperature
T10		223.6 deg. F	Steam temperature to second effect
T11		132.8 deg. F	Condenser condensate temperature
T12		159.0 deg. F	Separator vapor temperature
T15		277.0 deg. F	Steam temperature to first effect
T19	h1(T1)	222.0 deg. F	Solution temperature in first effect
T28		196.9 deg. F	Second effect steam condensate temperature
T29		55.7 deg. F	Cooling water inlet temperature
T34		158.0 deg. F	Product temperature

Note: Refer to Figures 2.1 and 2.2 for location of measuring devices.

EXPERIMENT OL8

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.00	5.00
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	3.34	3.33
FIRST EFFECT OVERHEAD FLOW	F5	1.71	1.65
PRODUCT FLOW	F6	1.67	1.58
SECOND EFFECT OVERHEAD FLOW	F7	1.49	1.55
CONDENSER COOLING WATER FLOW	F9	39.93	41.72
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.034	0.025
PRODUCT CONCENTRATION	C6	0.100	0.075
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.4	189.6
STEAM TO FIRST EFFECT	T15	280.1	279.6
SOLUTION IN FIRST EFFECT	T19	224.6	224.0
VAPOR IN FIRST EFFECT	T2	222.9	221.9
FIRST EFFECT STEAM CONDENSATE	T5	248.2	247.0
SOLUTION TO SECOND EFFECT	T4	181.8	181.4
STEAM TO SECOND EFFECT	T10	223.6	222.5
PRODUCT	T34	157.2	156.9
STEAM CONDENSATE SECOND EFFECT	T28	196.9	196.8
SEPARATOR VAPOR	T12	159.0	159.0
COOLING WATER INLET	T29	55.7	55.1
COOLING WATER OUTLET	T1	96.1	96.2
CONDENSER CONDENSATE	T11	132.8	139.6
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		5.65	5.64
SECOND EFFECT PRESSURE		-16.45	-16.47
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		1.92	3.12
TOTAL COMPONENT BALANCE		4.22	2.03

Table A-2 Steady State Data for Run OL8 (-30% step in CF)

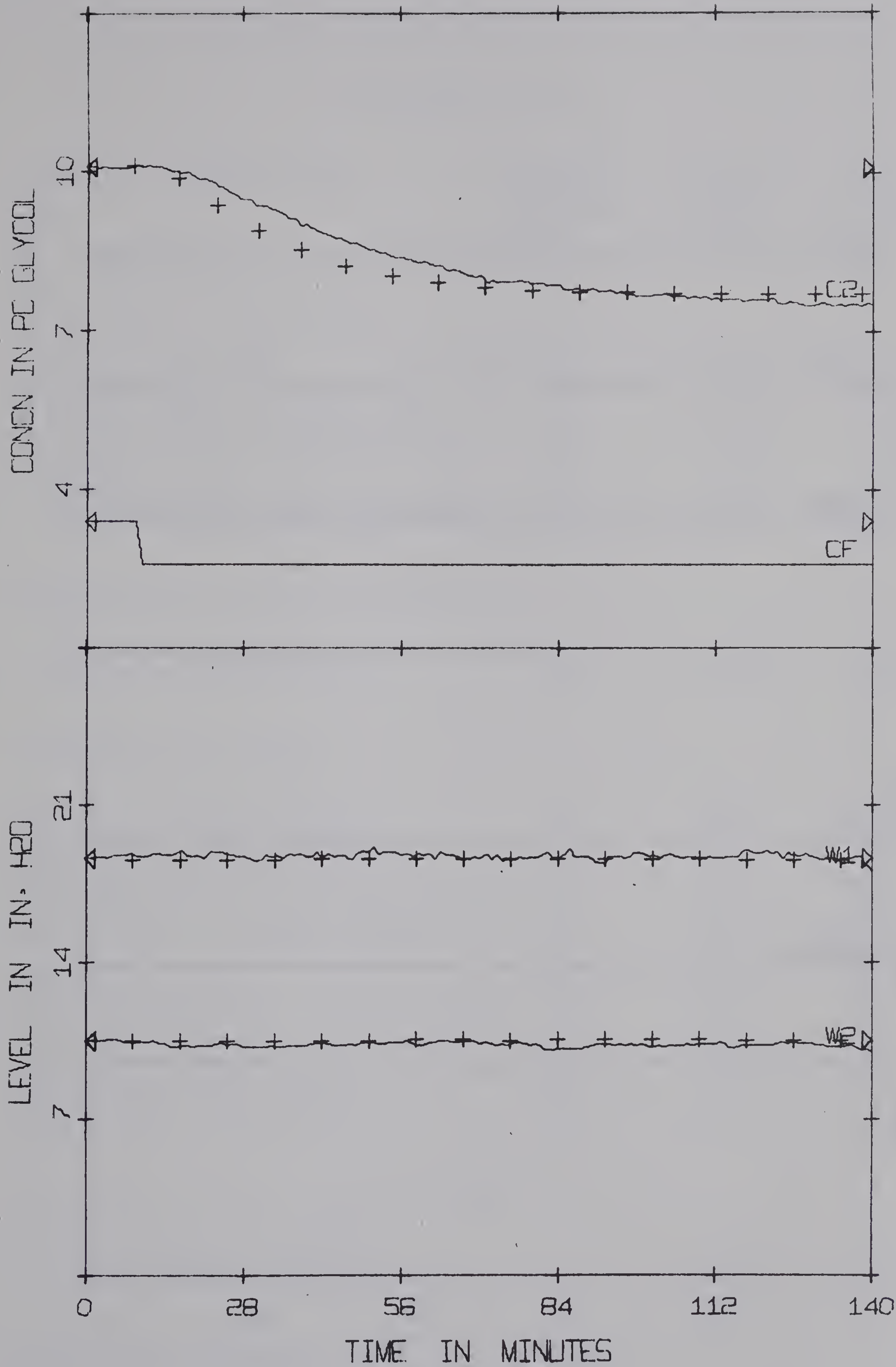
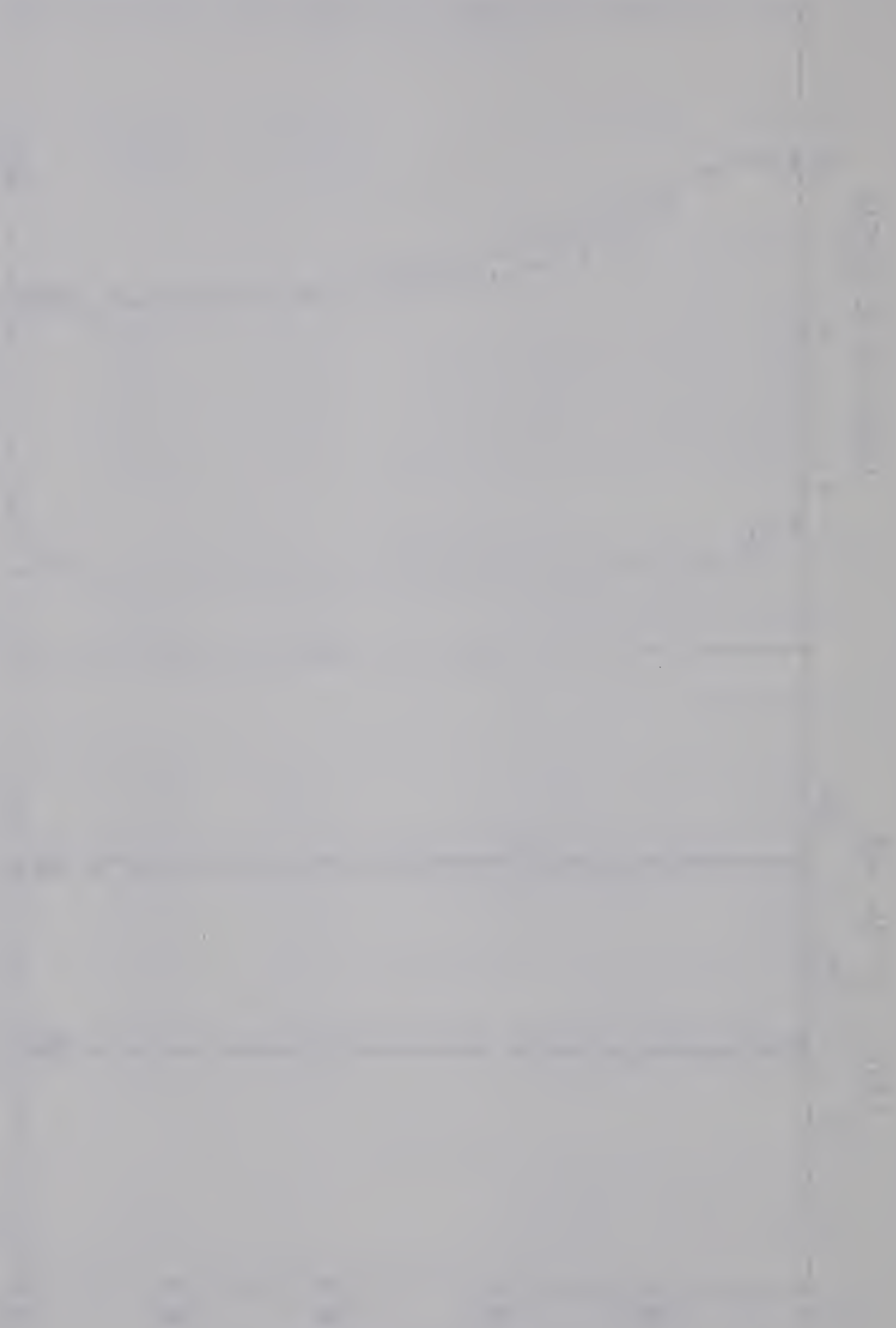


Figure A-1a Transient Data for Run OL8 (-30% step in CF)



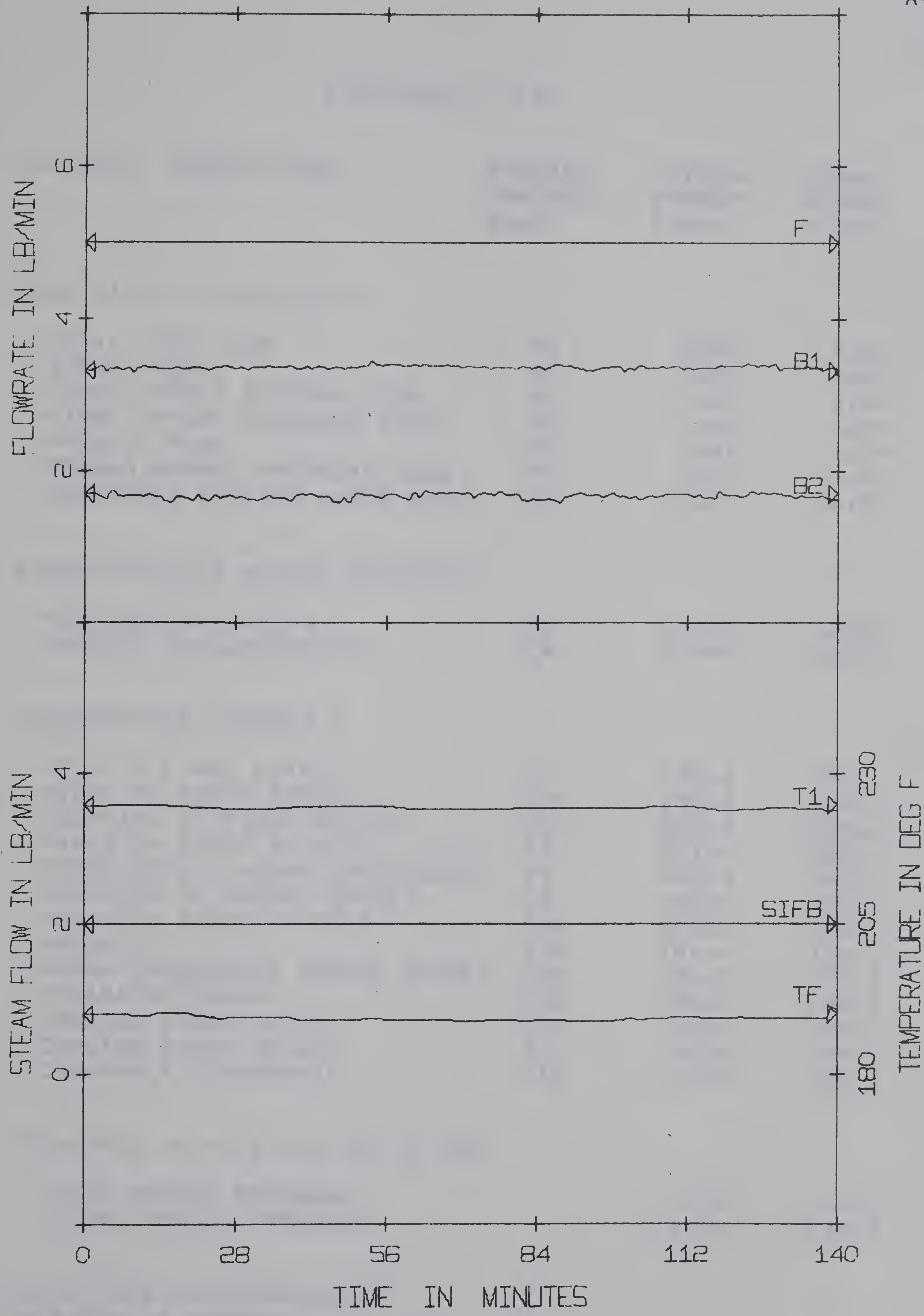


Figure A-1b Transient Data for Run OL8 (-30% step in CF)

EXPERIMENT OL9

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.08	5.60
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	3.27	3.83
FIRST EFFECT OVERHEAD FLOW	F5	1.66	1.66
PRODUCT FLOW	F6	1.58	2.14
SECOND EFFECT OVERHEAD FLOW	F7	2.26	1.70
CONDENSER COOLING WATER FLOW	F9	41.13	40.50
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.091	0.074
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.8	189.3
STEAM TO FIRST EFFECT	T15	279.2	278.6
SOLUTION IN FIRST EFFECT	T19	222.9	222.3
VAPOR IN FIRST EFFECT	T2	221.4	220.7
FIRST EFFECT STEAM CONDENSATE	T5	248.6	247.2
SOLUTION TO SECOND EFFECT	T4	181.0	182.0
STEAM TO SECOND EFFECT	T10	222.2	221.4
PRODUCT	T34	156.4	158.7
STEAM CONDENSATE SECOND EFFECT	T28	196.9	197.8
SEPARATOR VAPOR	T12	158.6	160.2
COOLING WATER INLET	T29	54.6	54.0
COOLING WATER OUTLET	T1	95.3	94.9
CONDENSER CONDENSATE	T11	147.8	137.3
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		4.92	5.25
SECOND EFFECT PRESSURE		-16.56	-16.52
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		-4.79	4.76
TOTAL COMPONENT BALANCE		4.86	4.78

Table A-3 Steady State Data for Run OL9 (+10% step in F)

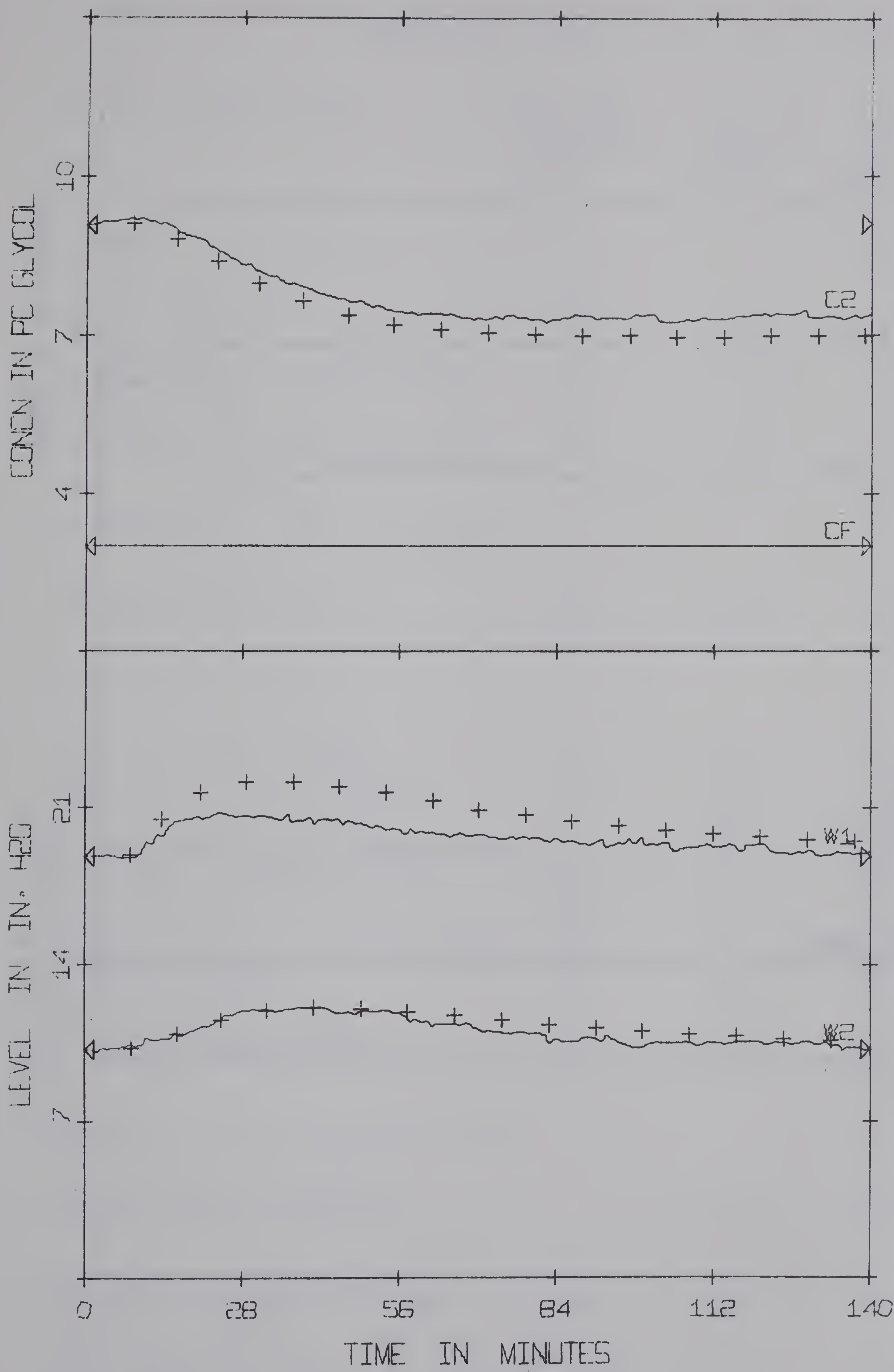


Figure A-2a Transient Data for Run OL9 (+10% step in F)

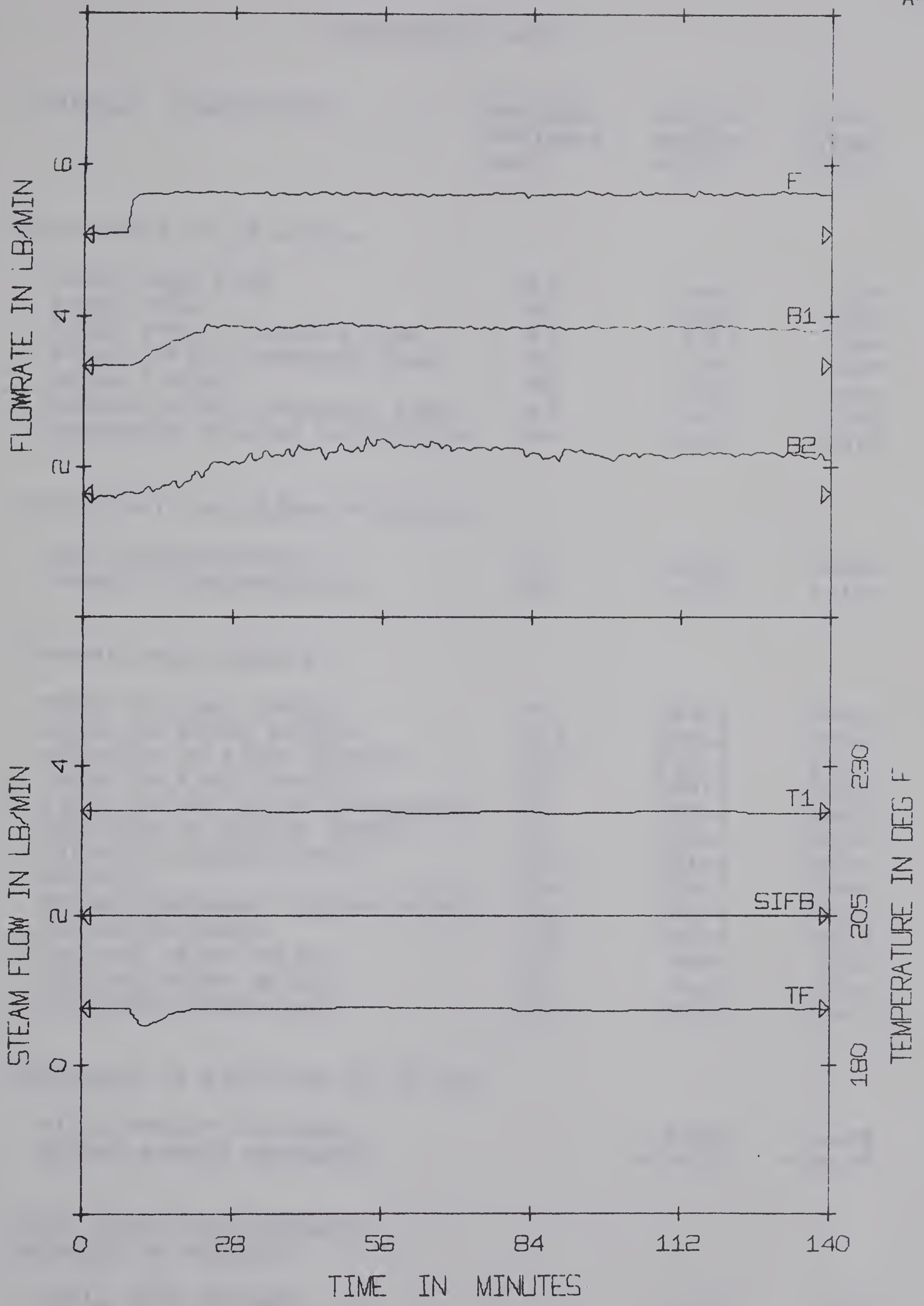
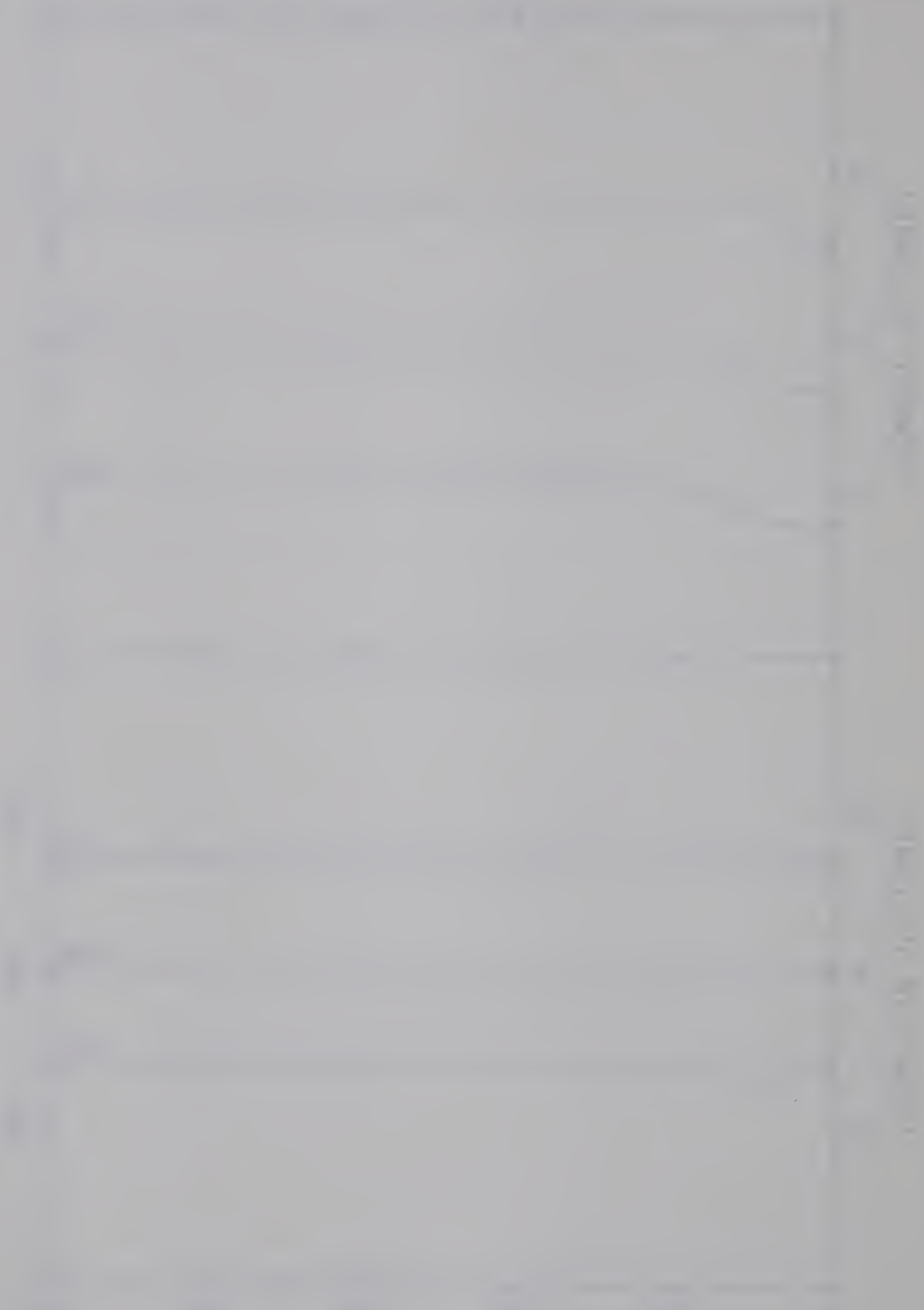


Figure A-2b Transient Data for Run OL9 (+10% step in F)



EXPERIMENT OL10

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.60	4.59
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	3.83	2.84
FIRST EFFECT OVERHEAD FLOW	F5	1.66	1.69
PRODUCT FLOW	F6	2.14	1.10
SECOND EFFECT OVERHEAD FLOW	F7	1.70	1.51
CONDENSER COOLING WATER FLOW	F9	40.50	41.28
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.074	0.120
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.3	190.3
STEAM TO FIRST EFFECT	T15	278.6	278.7
SOLUTION IN FIRST EFFECT	T19	222.3	221.8
VAPOR IN FIRST EFFECT	T2	220.7	220.2
FIRST EFFECT STEAM CONDENSATE	T5	247.2	246.2
SOLUTION TO SECOND EFFECT	T4	182.0	181.0
STEAM TO SECOND EFFECT	T10	221.4	221.0
PRODUCT	T34	158.7	155.6
STEAM CONDENSATE SECOND EFFECT	T28	197.8	197.3
SEPARATOR VAPOR	T12	160.2	159.8
COOLING WATER INLET	T29	54.0	54.0
COOLING WATER OUTLET	T1	94.9	93.3
CONDENSER CONDENSATE	T11	137.3	144.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		5.25	4.98
SECOND EFFECT PRESSURE		-16.52	-16.45
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.76	3.30
TOTAL COMPONENT BALANCE		4.78	4.40

Table A-4 Steady State Data for Run OL10 (-20% step in F)

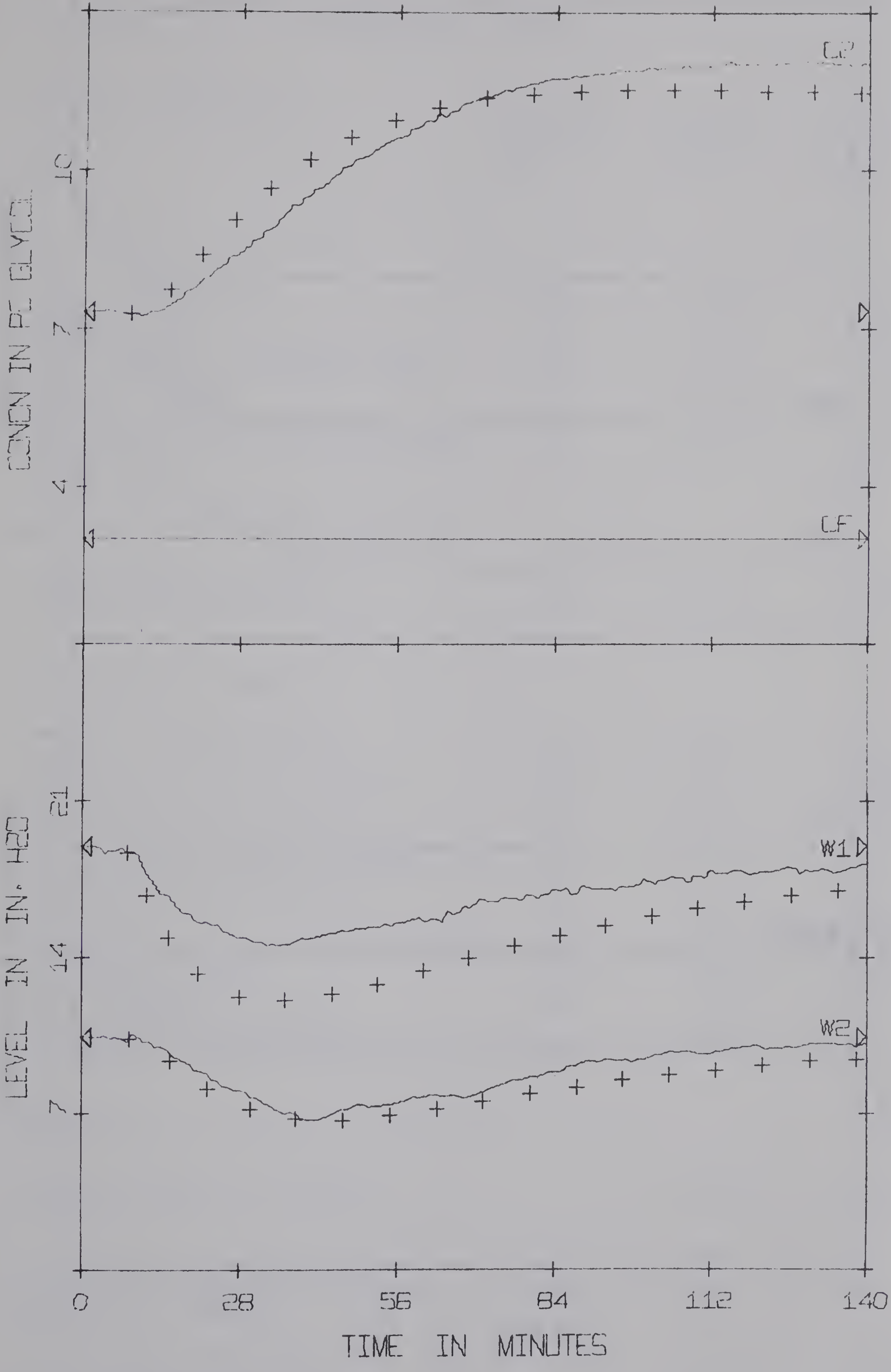


Figure A-3a Transient Data for Run OL10 (-20% step in F)

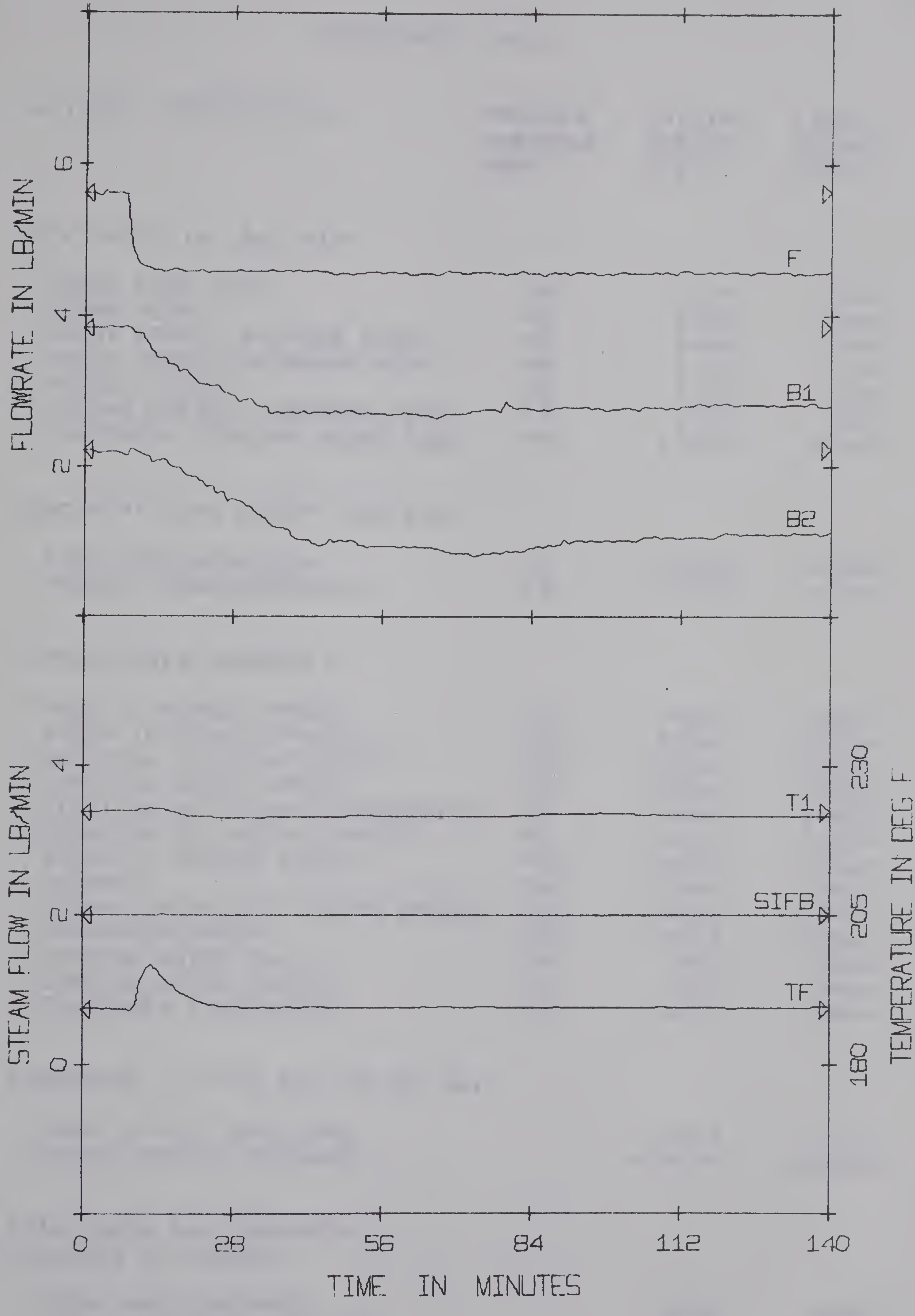
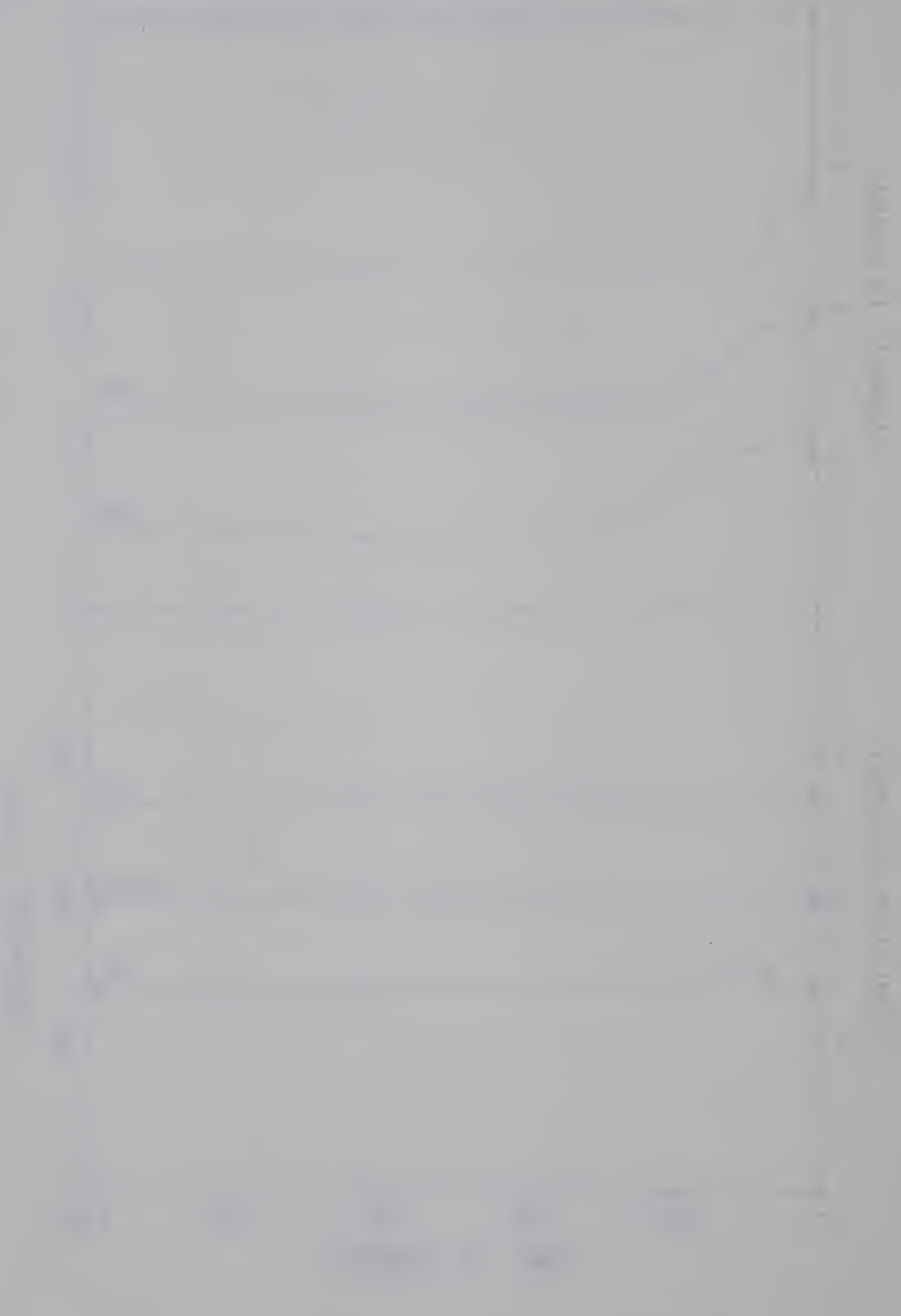


Figure A-3b Transient Data for Run OL10 (-20% step in F)



EXPERIMENT OL11

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.59	5.61
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	2.84	3.84
FIRST EFFECT OVERHEAD FLOW	F5	1.69	1.65
PRODUCT FLOW	F6	1.10	2.16
SECOND EFFECT OVERHEAD FLOW	F7	1.51	2.07
CONDENSER COOLING WATER FLOW	F9	41.28	40.82
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.120	0.079
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.3	189.1
STEAM TO FIRST EFFECT	T15	278.7	278.4
SOLUTION IN FIRST EFFECT	T19	221.8	222.2
VAPOR IN FIRST EFFECT	T2	220.2	220.5
FIRST EFFECT STEAM CONDENSATE	T5	246.2	246.3
SOLUTION TO SECOND EFFECT	T4	181.0	182.4
STEAM TO SECOND EFFECT	T10	221.0	221.0
PRODUCT	T34	155.6	158.2
STEAM CONDENSATE SECOND EFFECT	T28	197.3	197.0
SEPARATOR VAPOR	T12	159.8	159.5
COOLING WATER INLET	T29	54.0	53.8
COOLING WATER OUTLET	T1	93.3	94.5
CONDENSER CONDENSATE	T11	144.1	139.4
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		4.98	5.02
SECOND EFFECT PRESSURE		-16.45	-16.56
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.30	3.03
TOTAL COMPONENT BALANCE		4.40	-0.62

Table A-5 Steady State Data for Run OL11 (+20% step in F)

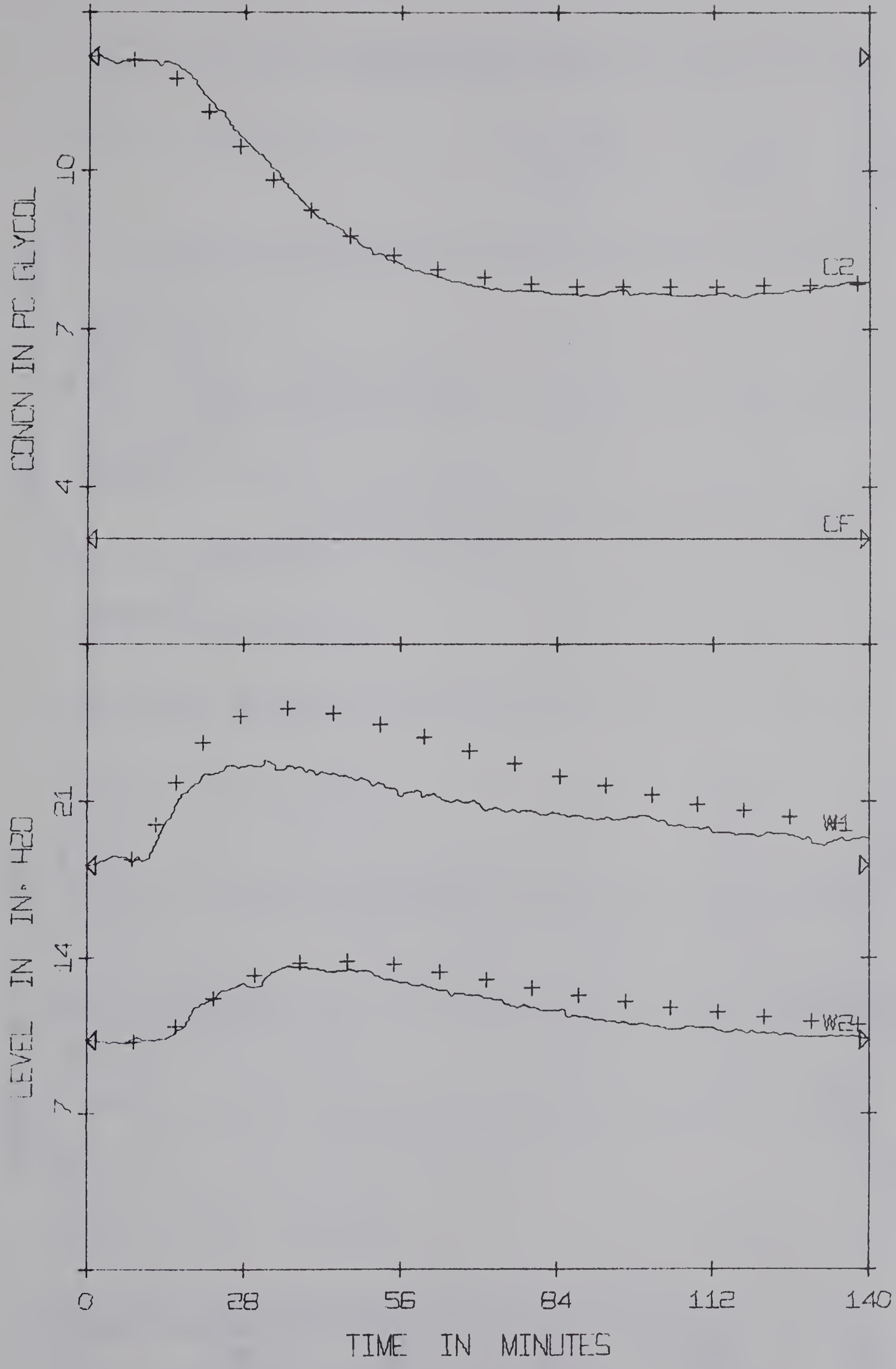


Figure A-4a Transient Data for Run OL11 (+20% step in F)

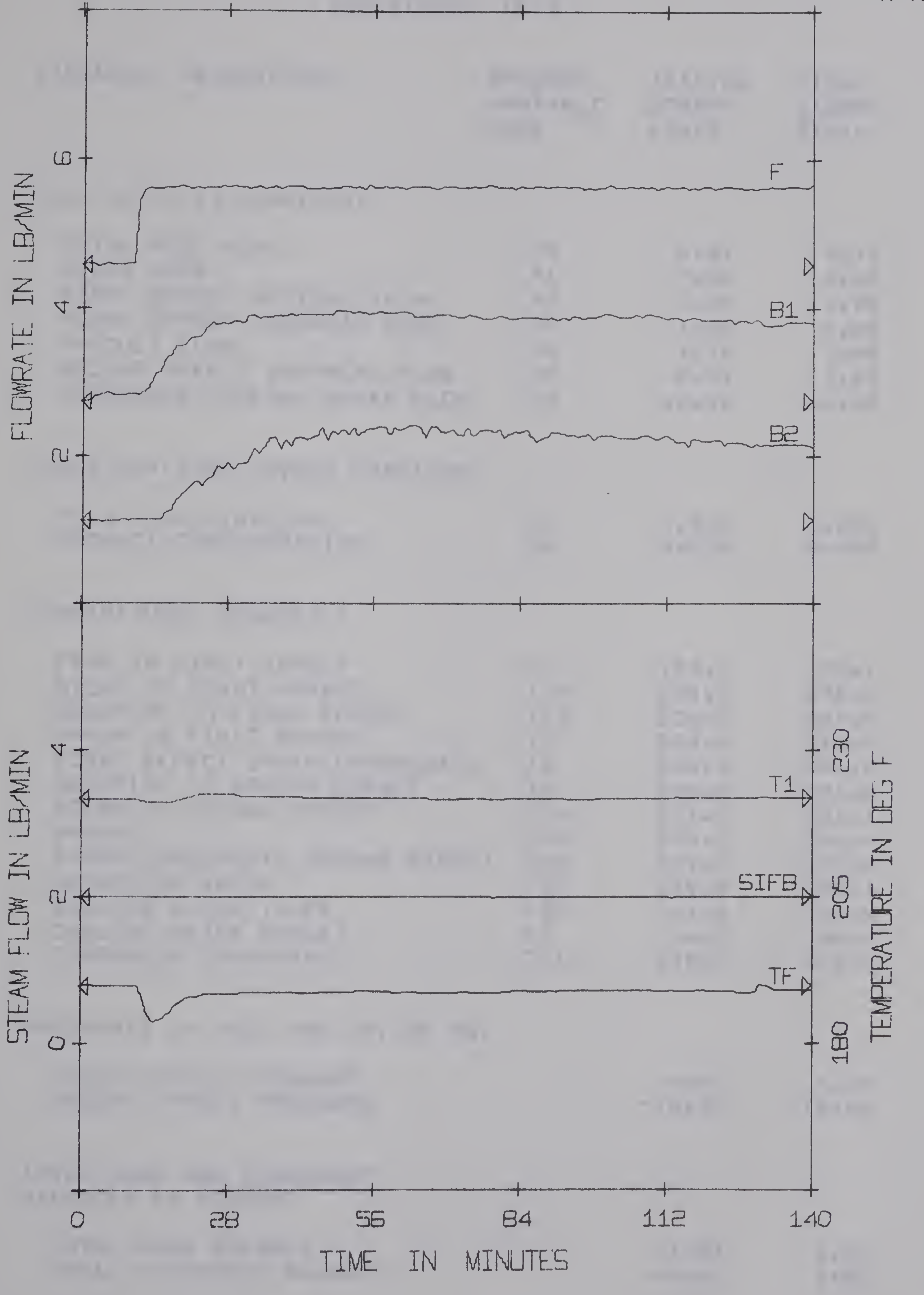


Figure A-4b Transient Data for Run OL11 (+20% step in F)

EXPERIMENT OL12

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.61	5.11
STEAM FLOW	F1	2.00	1.99
FIRST EFFECT BOTTOMS FLOW	F2	3.84	3.28
FIRST EFFECT OVERHEAD FLOW	F5	1.65	1.68
PRODUCT FLOW	F6	2.16	1.58
SECOND EFFECT OVERHEAD FLOW	F7	2.07	1.53
CONDENSER COOLING WATER FLOW	F9	40.82	41.09
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.079	0.093
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.1	190.1
STEAM TO FIRST EFFECT	T15	278.4	278.7
SOLUTION IN FIRST EFFECT	T19	222.2	221.9
VAPOR IN FIRST EFFECT	T2	220.5	220.3
FIRST EFFECT STEAM CONDENSATE	T5	246.3	246.2
SOLUTION TO SECOND EFFECT	T4	182.4	181.8
STEAM TO SECOND EFFECT	T10	221.0	221.1
PRODUCT	T34	158.2	156.4
STEAM CONDENSATE SECOND EFFECT	T28	197.0	197.0
SEPARATOR VAPOR	T12	159.5	158.7
COOLING WATER INLET	T29	53.8	54.4
COOLING WATER OUTLET	T1	94.5	93.9
CONDENSER CONDENSATE	T11	139.4	138.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		5.02	4.80
SECOND EFFECT PRESSURE		-16.56	-16.45
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.03	4.70
TOTAL COMPONENT BALANCE		-0.62	3.80

Table A-6 Steady State Data for Run OL12 (-10% step in F)

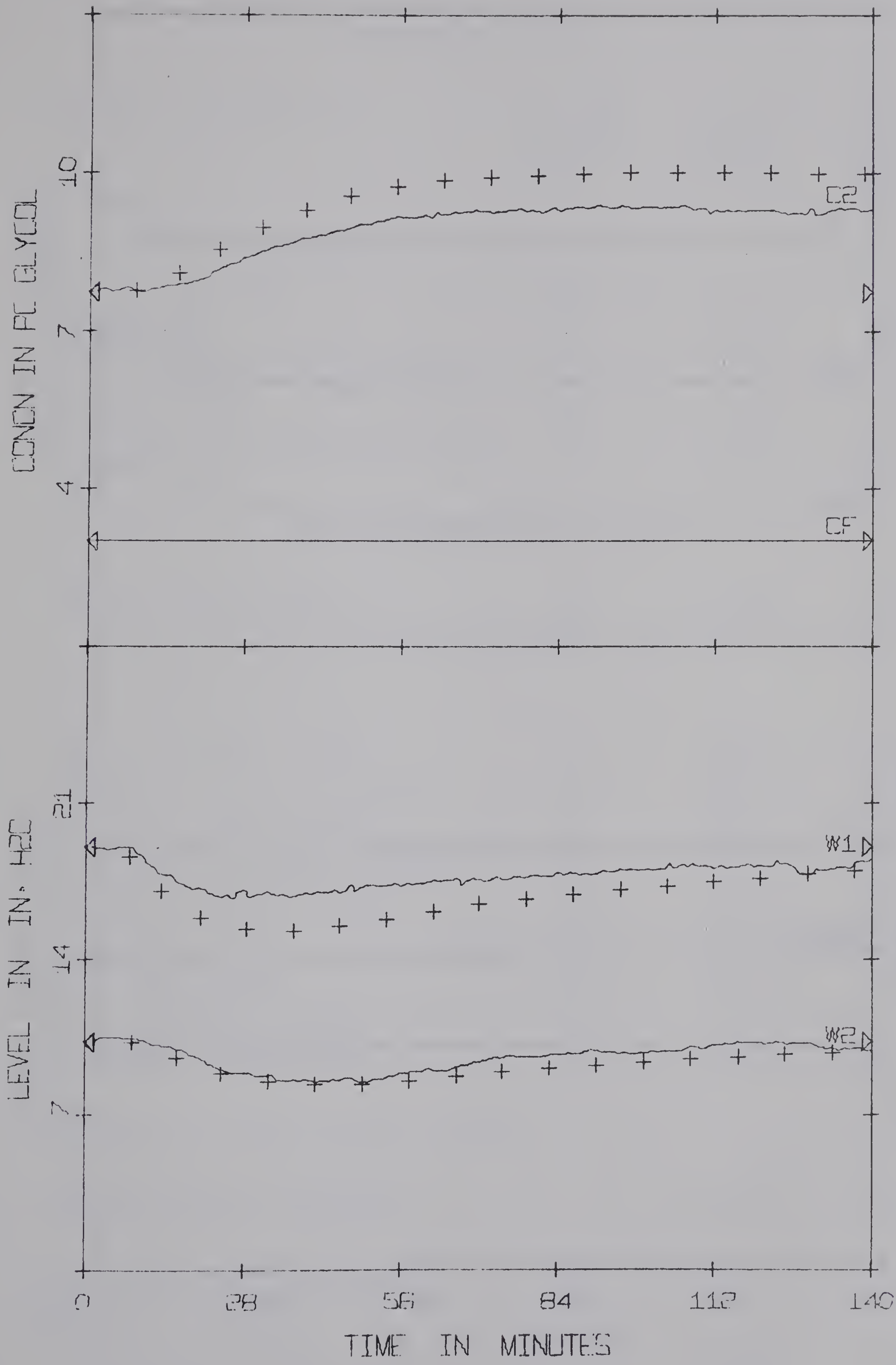


Figure A-5a Transient Data for Run OL12 (-10% step in F)

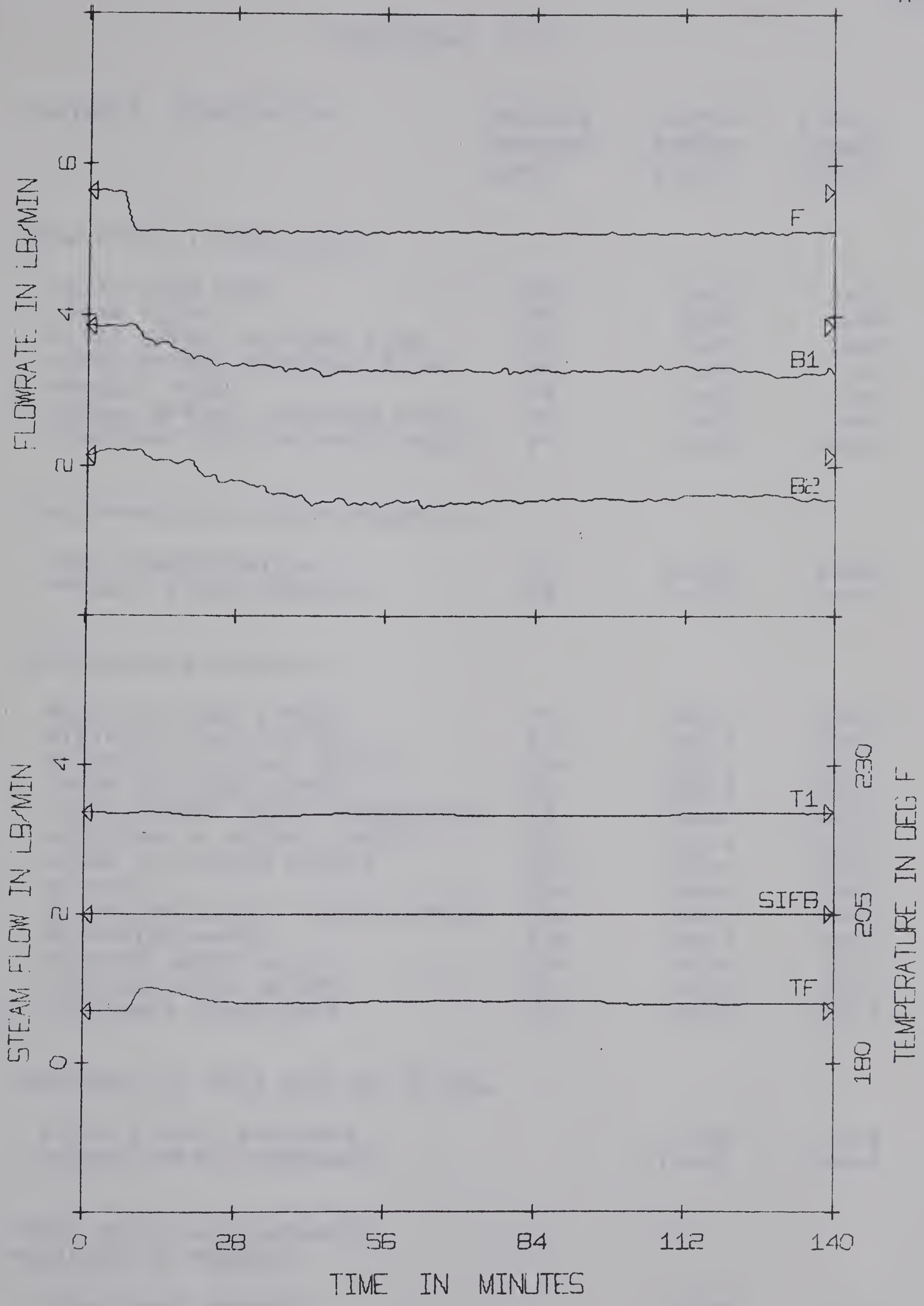


Figure A-5b Transient Data for Run OL12 (-10% step in F)

EXPERIMENT OL13

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.09	5.05
STEAM FLOW	F1	2.00	1.99
FIRST EFFECT BOTTOMS FLOW	F2	3.35	3.48
FIRST EFFECT OVERHEAD FLOW	F5	1.67	1.49
PRODUCT FLOW	F6	1.59	1.95
SECOND EFFECT OVERHEAD FLOW	F7	1.52	1.40
CONDENSER COOLING WATER FLOW	F9	41.32	40.81
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.093	0.076
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.6	156.9
STEAM TO FIRST EFFECT	T15	278.7	278.7
SOLUTION IN FIRST EFFECT	T19	221.9	221.7
VAPOR IN FIRST EFFECT	T2	220.3	220.2
FIRST EFFECT STEAM CONDENSATE	T5	246.0	245.9
SOLUTION TO SECOND EFFECT	T4	181.5	181.2
STEAM TO SECOND EFFECT	T10	221.0	220.9
PRODUCT	T34	156.6	156.6
STEAM CONDENSATE SECOND EFFECT	T28	196.2	195.3
SEPARATOR VAPOR	T12	158.7	158.4
COOLING WATER INLET	T29	54.4	54.1
COOLING WATER OUTLET	T1	94.8	91.7
CONDENSER CONDENSATE	T11	145.3	140.5
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		4.79	4.65
SECOND EFFECT PRESSURE		-16.45	-16.47
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.28	3.76
TOTAL COMPONENT BALANCE		1.95	3.39

Table A-7 Steady State Data for Run OL13 (-18% step in TF)

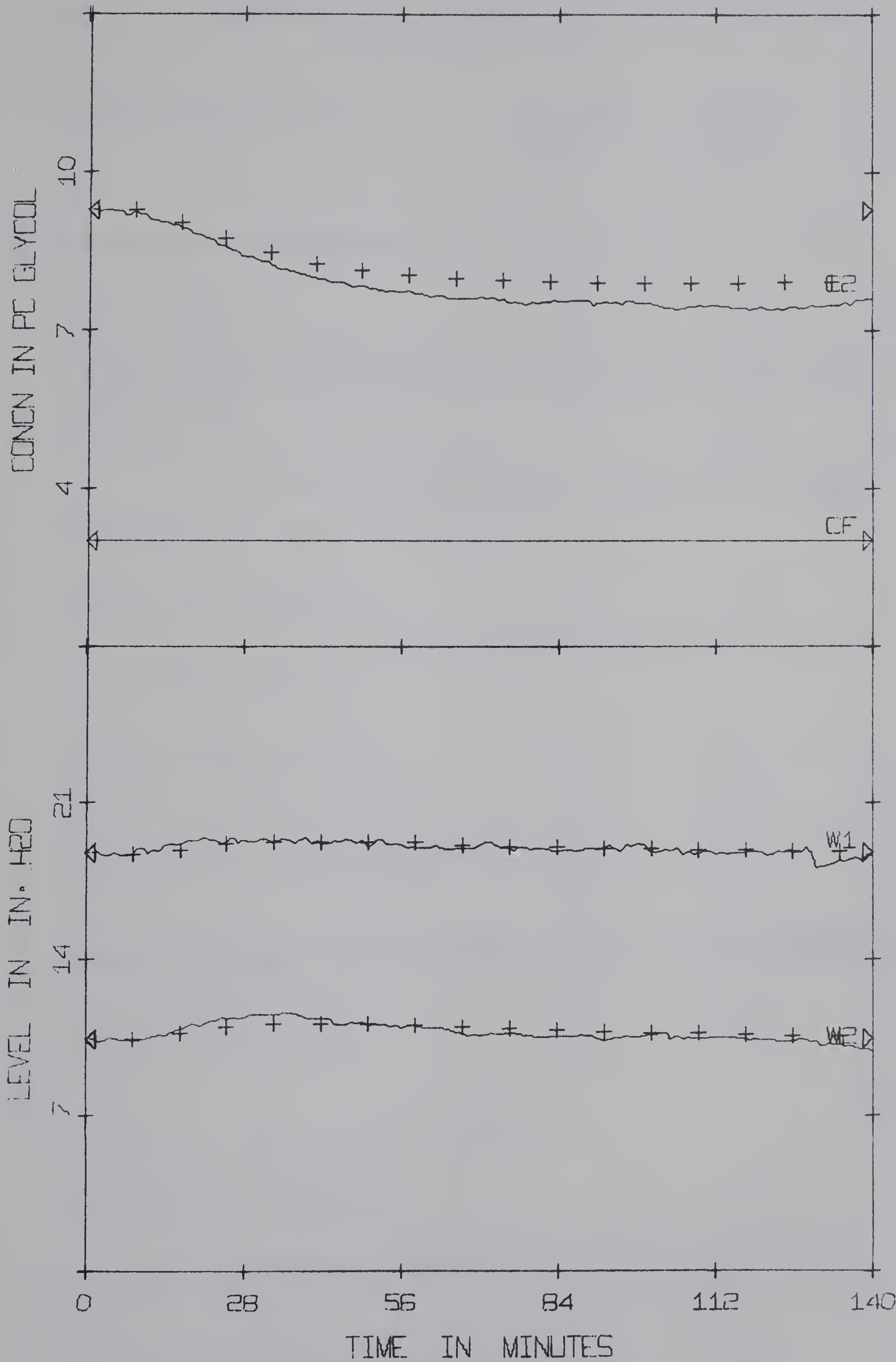


Figure A-6a Transient Data for Run OL13 (-18% step in TF)

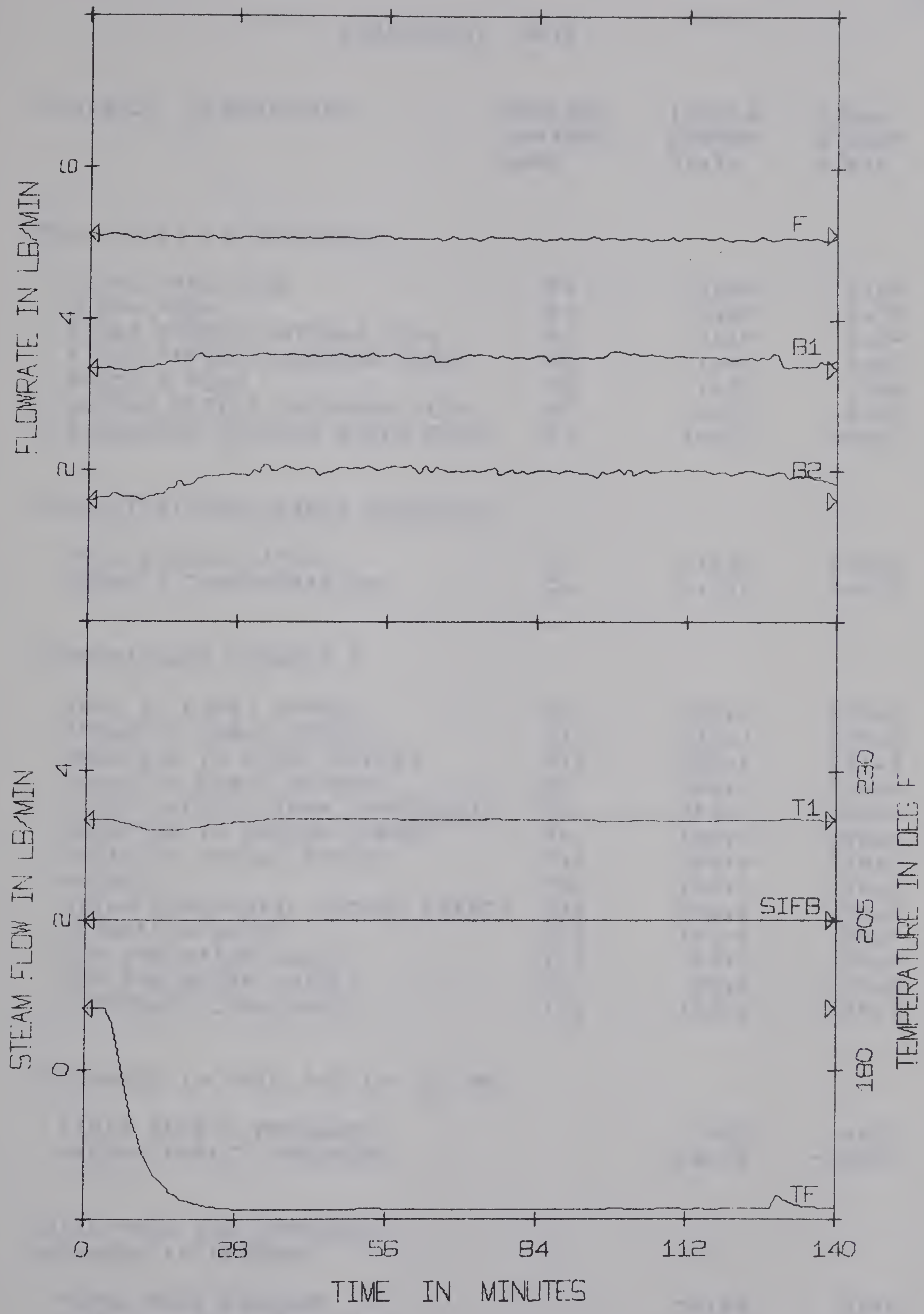


Figure A-6b Transient Data for Run OL13 (-18% step in TF)

EXPERIMENT OL15

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.08	5.09
STEAM FLOW	F1	2.20	1.79
FIRST EFFECT BOTTOMS FLOW	F2	3.18	3.44
FIRST EFFECT OVERHEAD FLOW	F5	1.81	1.51
PRODUCT FLOW	F6	1.37	1.96
SECOND EFFECT OVERHEAD FLOW	F7	1.73	1.47
CONDENSER COOLING WATER FLOW	F9	40.15	39.91
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.110	0.079
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.2	190.0
STEAM TO FIRST EFFECT	T15	277.7	279.2
SOLUTION IN FIRST EFFECT	T19	225.8	221.3
VAPOR IN FIRST EFFECT	T2	224.1	219.6
FIRST EFFECT STEAM CONDENSATE	T5	251.4	243.1
SOLUTION TO SECOND EFFECT	T4	182.5	182.5
STEAM TO SECOND EFFECT	T10	224.6	220.2
PRODUCT	T34	158.1	158.0
STEAM CONDENSATE SECOND EFFECT	T28	201.6	196.1
SEPARATOR VAPOR	T12	161.0	159.7
COOLING WATER INLET	T29	53.9	53.9
COOLING WATER OUTLET	T1	99.0	93.6
CONDENSER CONDENSATE	T11	152.2	146.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		7.29	4.70
SECOND EFFECT PRESSURE		-16.43	-16.43
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		-1.13	5.78
TOTAL COMPONENT BALANCE		3.30	0.65

Table A-8 Steady State Data for Run OL15 (-20% step in SIFB)

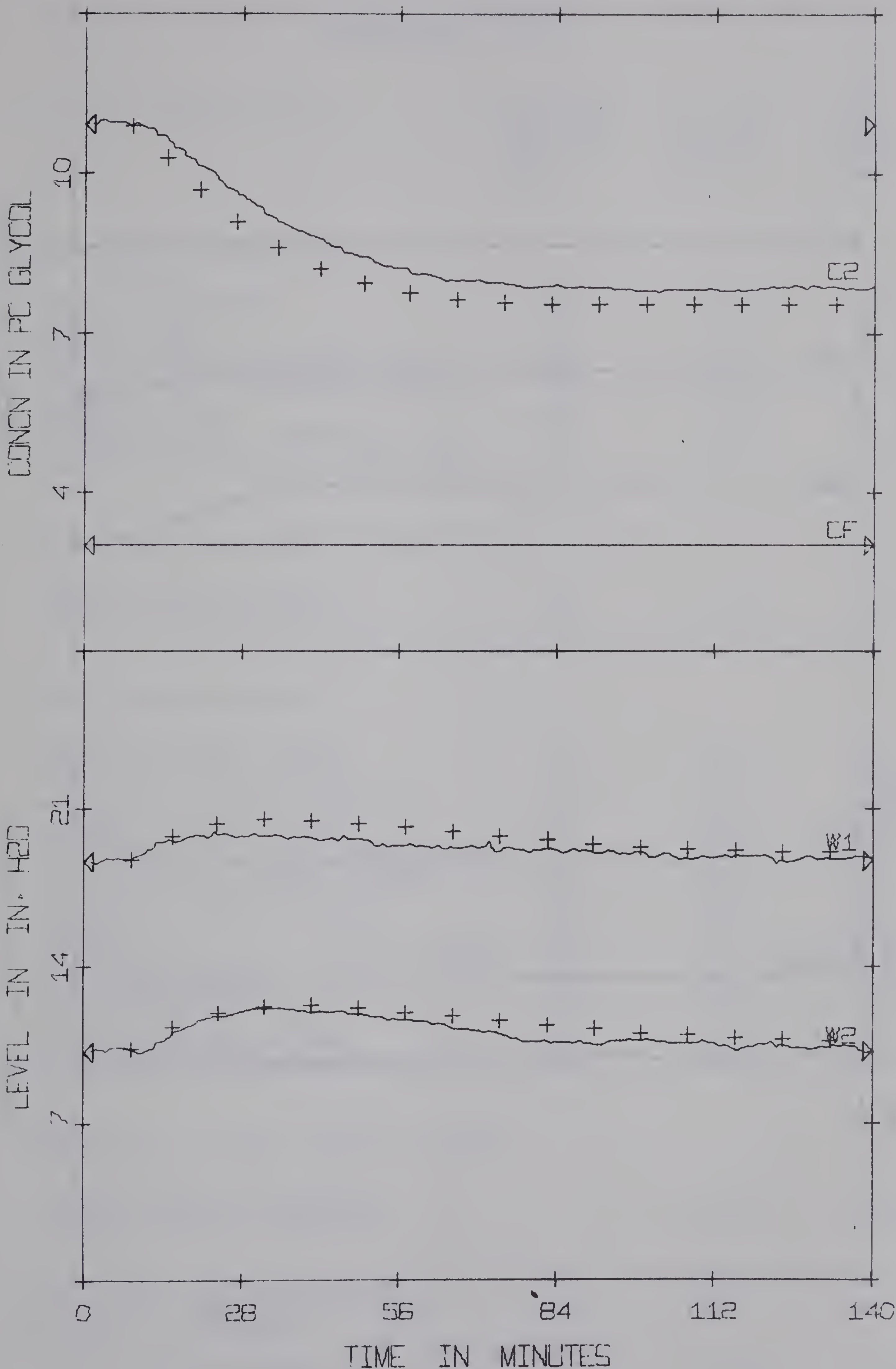


Figure A-7a Transient Data for Run OL15 (-20% step in SIFB)

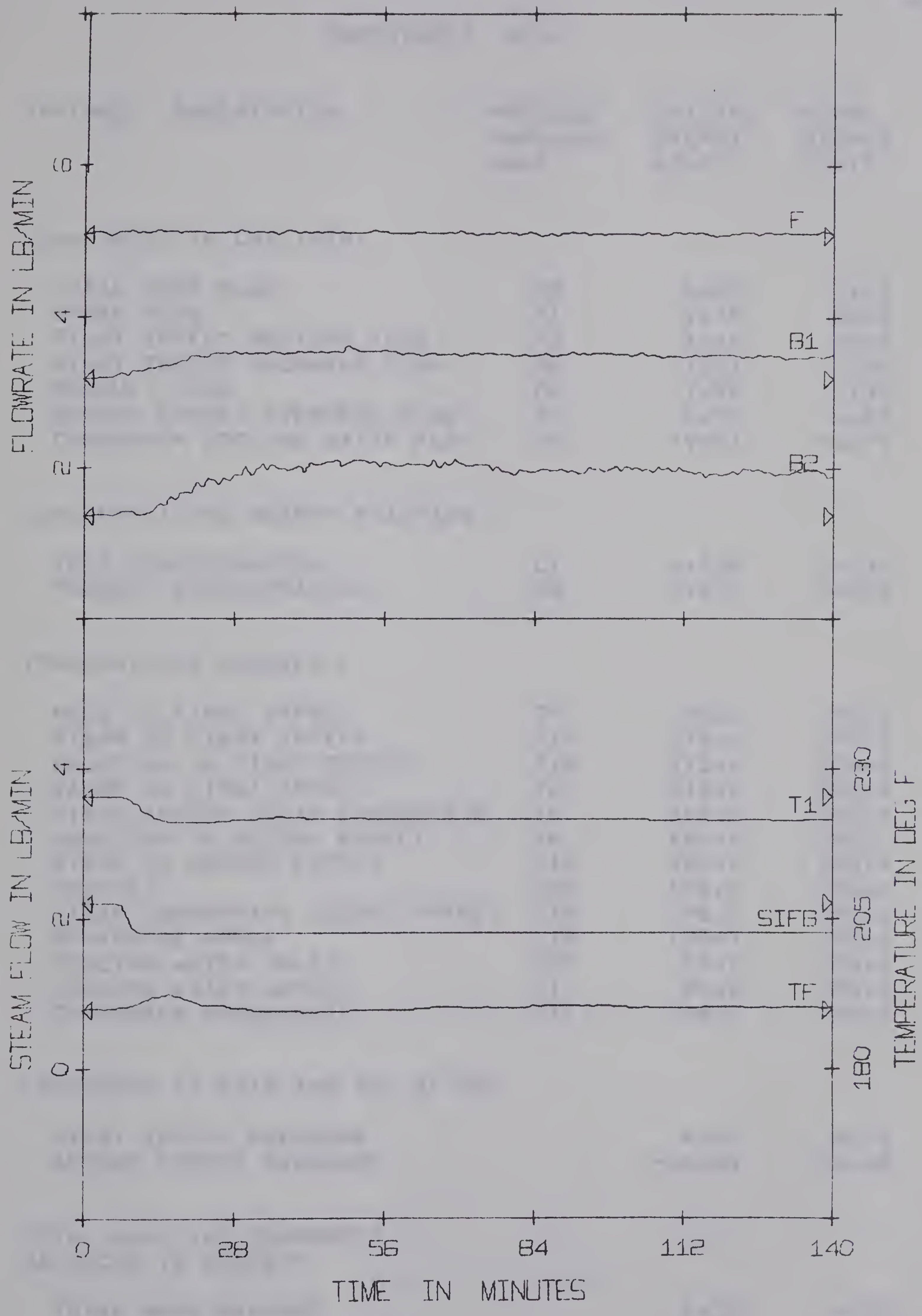


Figure A-7b Transient Data for Run OL15 (-20% step in SIFB)

EXPERIMENT OL16

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.09	5.10
STEAM FLOW	F1	1.79	2.19
FIRST EFFECT BOTTOMS FLOW	F2	3.44	3.18
FIRST EFFECT OVERHEAD FLOW	F5	1.51	1.80
PRODUCT FLOW	F6	1.96	1.34
SECOND EFFECT OVERHEAD FLOW	F7	1.47	1.93
CONDENSER COOLING WATER FLOW	F9	39.91	40.75
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.079	0.112
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.0	190.4
STEAM TO FIRST EFFECT	T15	279.2	277.7
SOLUTION IN FIRST EFFECT	T19	221.3	224.8
VAPOR IN FIRST EFFECT	T2	219.6	222.9
FIRST EFFECT STEAM CONDENSATE	T5	243.1	250.3
SOLUTION TO SECOND EFFECT	T4	182.5	181.1
STEAM TO SECOND EFFECT	T10	220.2	223.4
PRODUCT	T34	158.0	156.4
STEAM CONDENSATE SECOND EFFECT	T28	196.1	199.1
SEPARATOR VAPOR	T12	159.7	159.0
COOLING WATER INLET	T29	53.9	54.1
COOLING WATER OUTLET	T1	93.6	98.2
CONDENSER CONDENSATE	T11	146.1	146.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		4.70	6.51
SECOND EFFECT PRESSURE		-16.43	-16.58
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		5.78	4.76
TOTAL COMPONENT BALANCE		0.65	1.75

Table A-9 Steady State Data for Run OL16 (+20% step in SIFB)

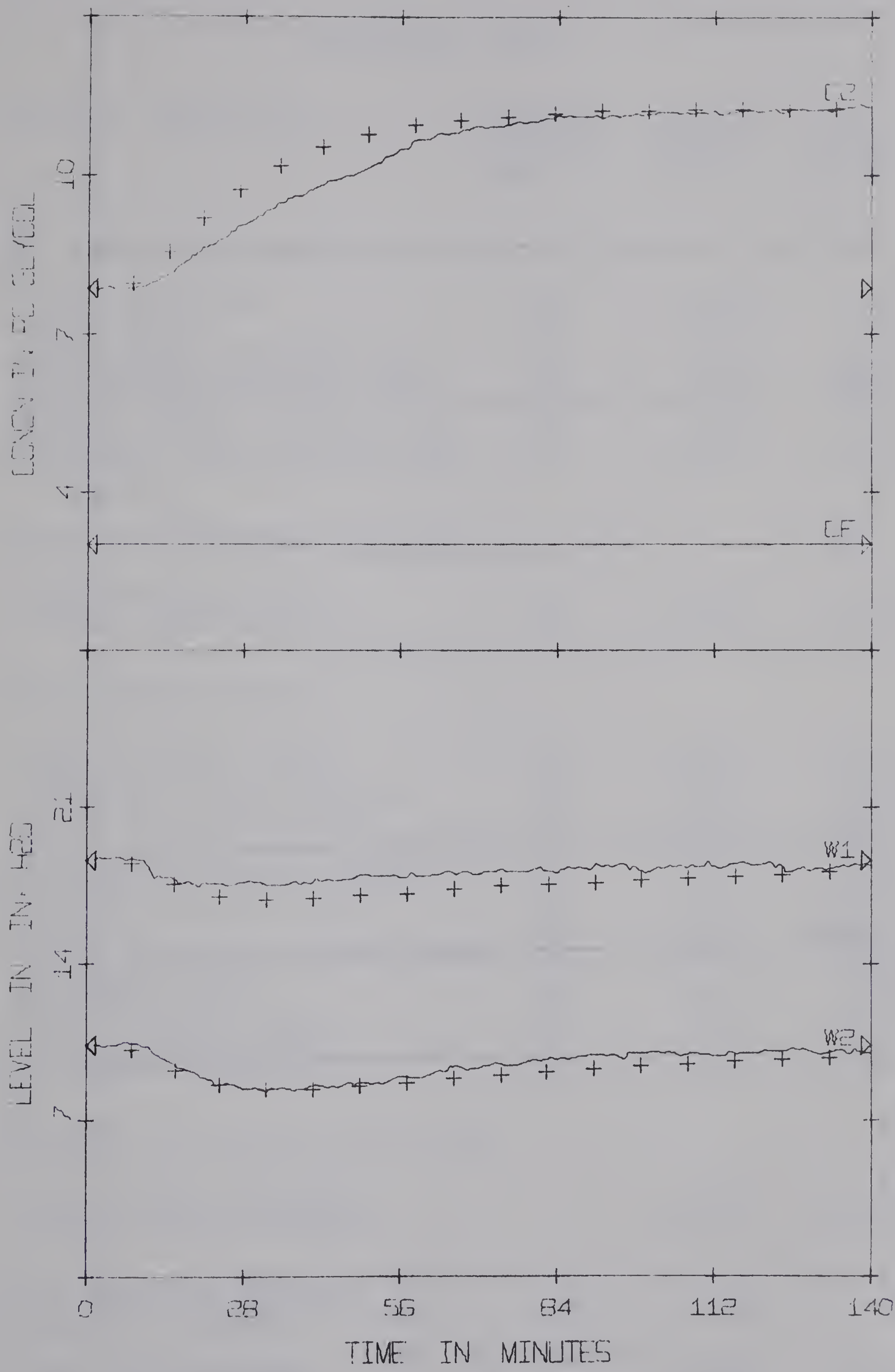
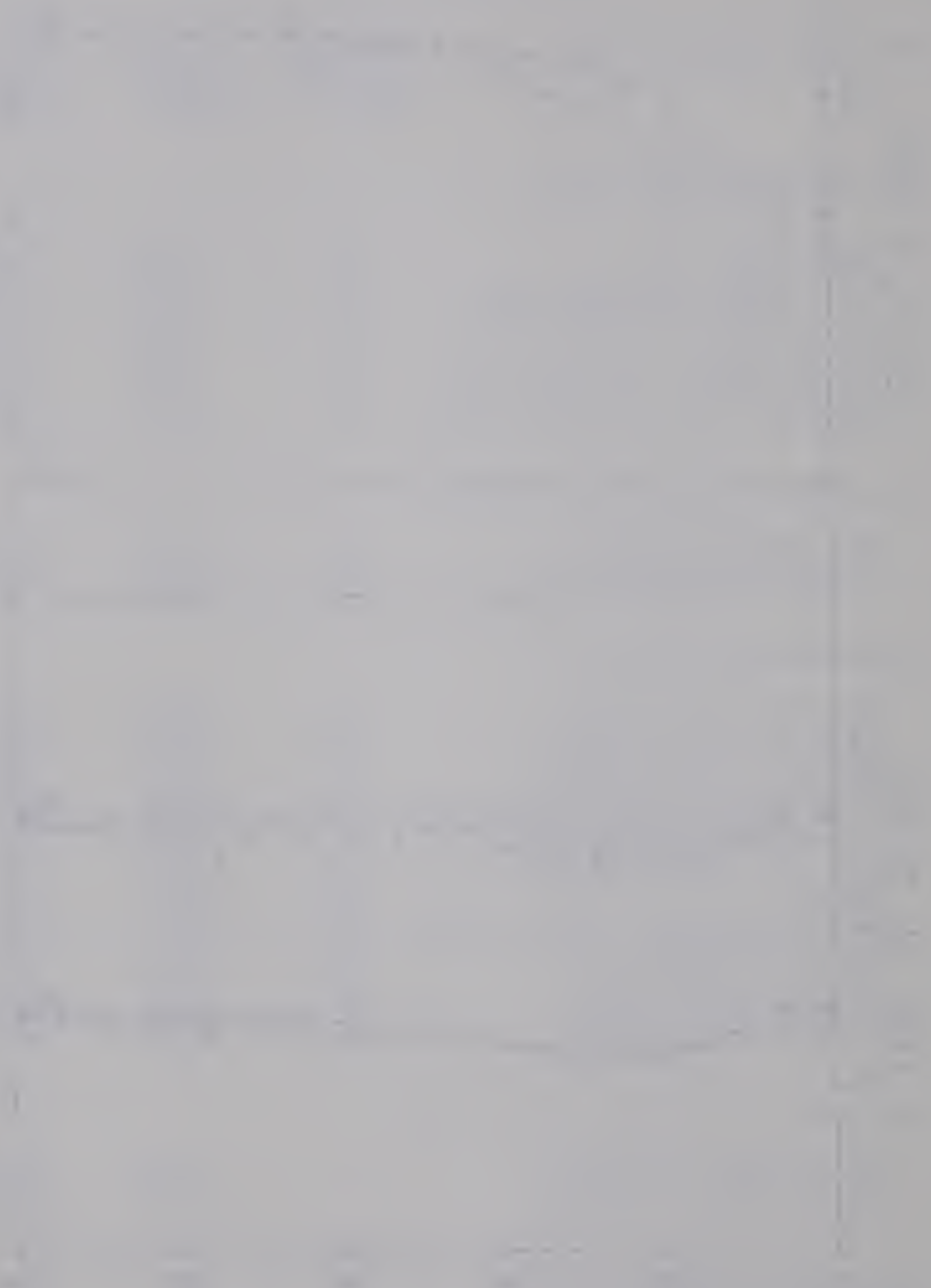


Figure A-8a Transient Data for Run OL16 (+20% step in SIFB)



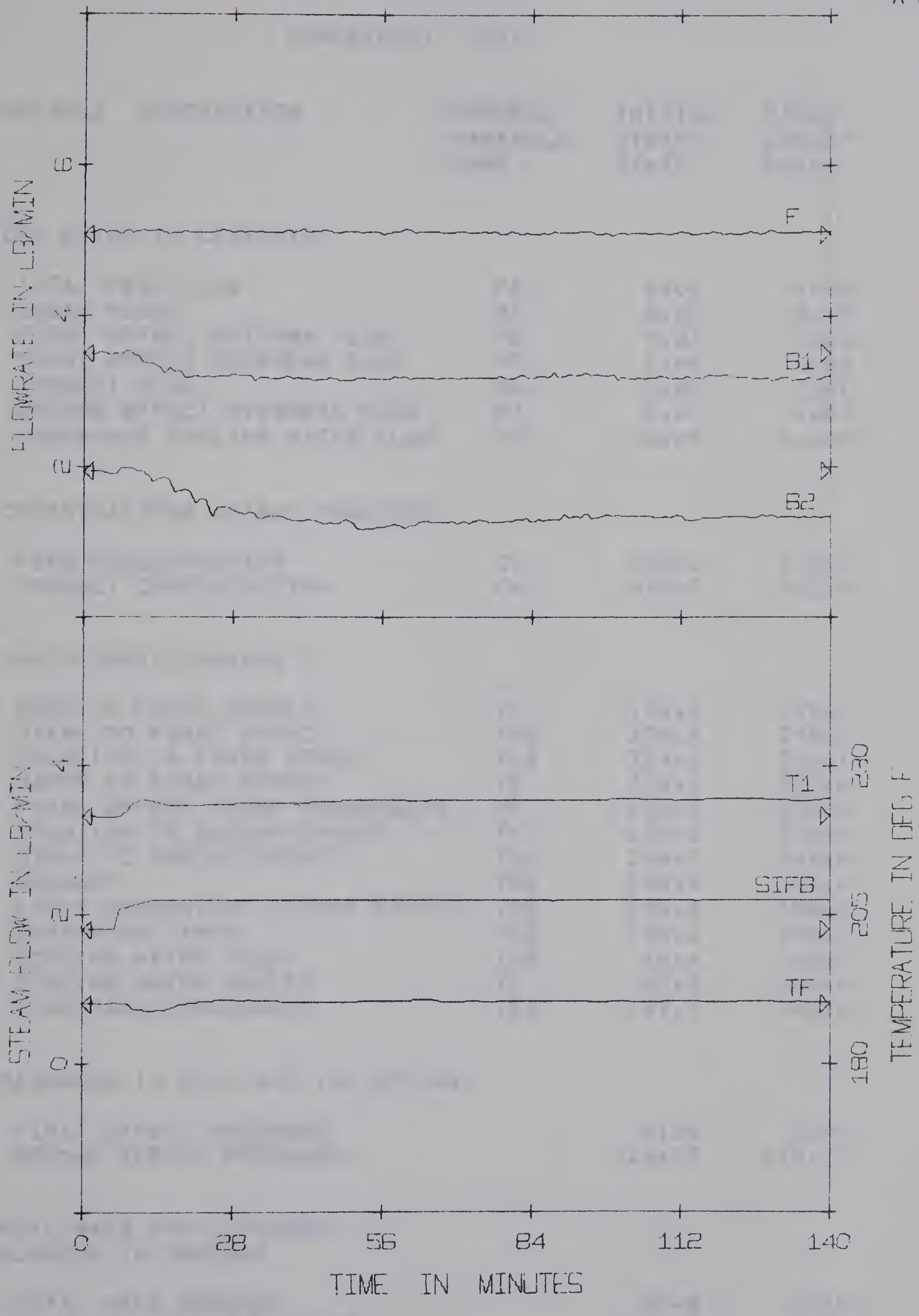


Figure A-8b Transient Data for Run OL16 (+20% step in SIFB)

EXPERIMENT OL18

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.03	4.95
STEAM FLOW	F1	2.00	2.19
FIRST EFFECT BOTTOMS FLOW	F2	3.21	3.04
FIRST EFFECT OVERHEAD FLOW	F5	1.68	1.82
PRODUCT FLOW	F6	1.59	1.21
SECOND EFFECT OVERHEAD FLOW	F7	1.57	1.67
CONDENSER COOLING WATER FLOW	F9	40.09	40.05
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.095	0.116
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.9	189.9
STEAM TO FIRST EFFECT	T15	279.3	278.7
SOLUTION IN FIRST EFFECT	T19	219.6	223.7
VAPOR IN FIRST EFFECT	T2	218.3	222.4
FIRST EFFECT STEAM CONDENSATE	T5	245.6	251.4
SOLUTION TO SECOND EFFECT	T4	179.1	178.8
STEAM TO SECOND EFFECT	T10	218.7	223.0
PRODUCT	T34	154.4	153.4
STEAM CONDENSATE SECOND EFFECT	T28	194.2	198.1
SEPARATOR VAPOR	T12	156.6	156.4
COOLING WATER INLET	T29	50.4	49.7
COOLING WATER OUTLET	T1	92.2	93.8
CONDENSER CONDENSATE	T11	137.6	143.2
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		4.52	6.94
SECOND EFFECT PRESSURE		-13.72	-13.72
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.48	5.34
TOTAL COMPONENT BALANCE		-1.58	6.18

Table A-10 Steady State Data for Run OL18 (+10% step in SIFB)

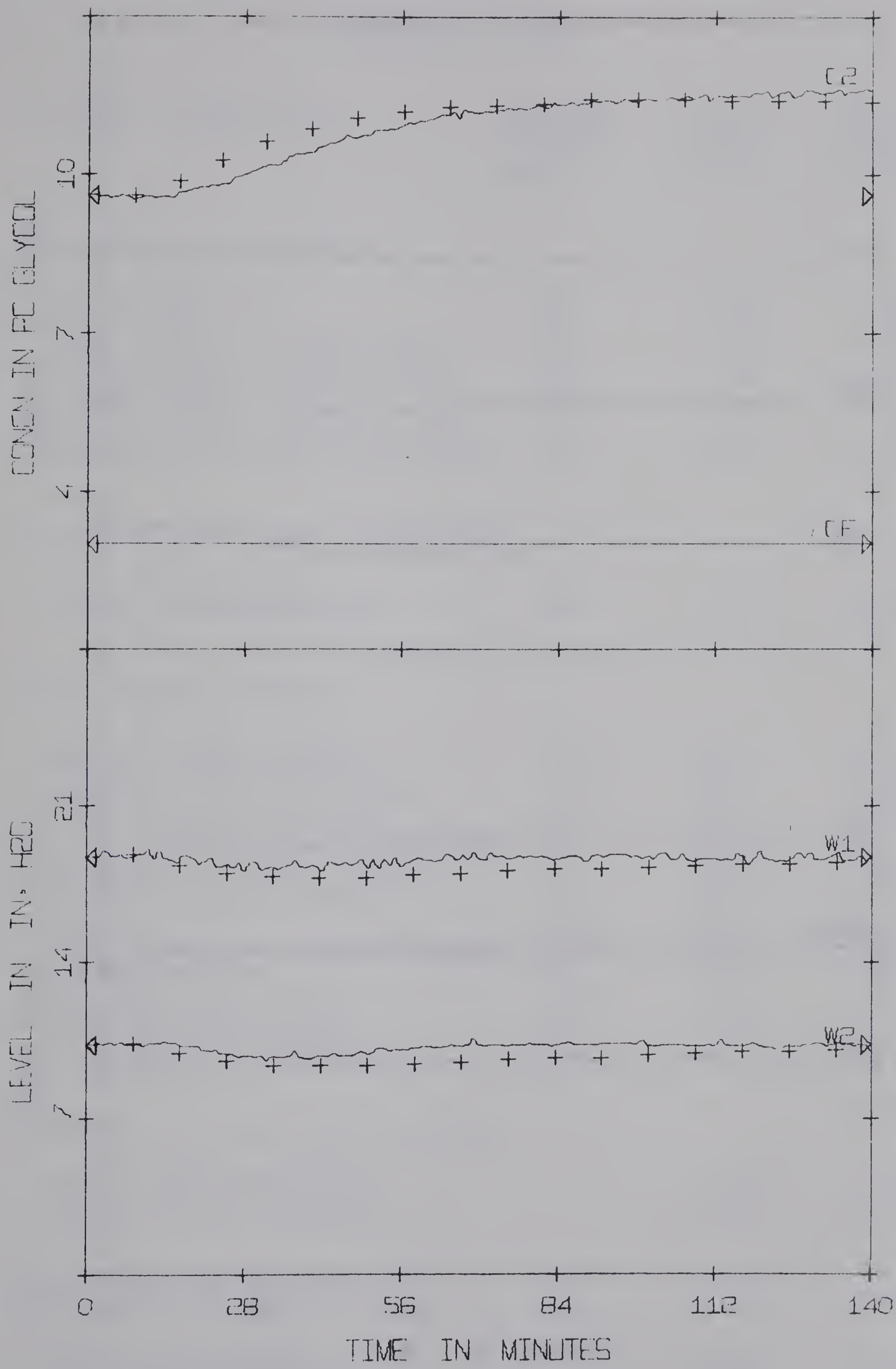


Figure A-9a Transient Data for Run OL18 (+10% step in SIFB)

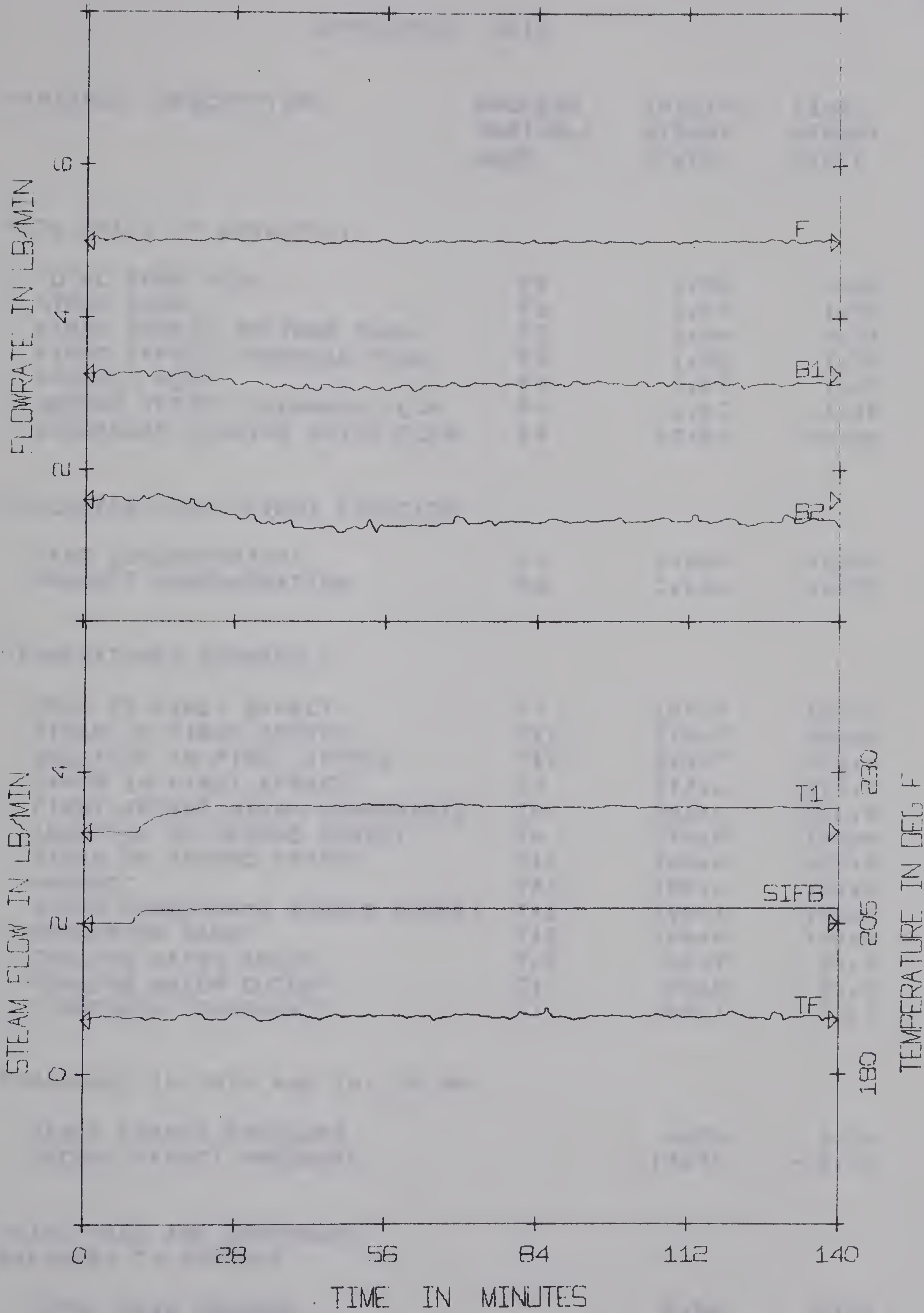


Figure A-9b Transient Data for Run OL18 (+10% step in SIFB)

EXPERIMENT OL19

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.95	5.03
STEAM FLOW	F1	2.19	1.99
FIRST EFFECT BOTTOMS FLOW	F2	3.04	3.21
FIRST EFFECT OVERHEAD FLOW	F5	1.82	1.70
PRODUCT FLOW	F6	1.21	1.47
SECOND EFFECT OVERHEAD FLOW	F7	1.67	1.56
CONDENSER COOLING WATER FLOW	F9	40.05	40.09
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.116	0.097
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.9	188.7
STEAM TO FIRST EFFECT	T15	278.7	280.2
SOLUTION IN FIRST EFFECT	T19	223.7	218.6
VAPOR IN FIRST EFFECT	T2	222.4	217.3
FIRST EFFECT STEAM CONDENSATE	T5	251.4	244.5
SOLUTION TO SECOND EFFECT	T4	178.8	178.4
STEAM TO SECOND EFFECT	T10	223.0	217.9
PRODUCT	T34	153.4	153.0
STEAM CONDENSATE SECOND EFFECT	T28	198.1	192.1
SEPARATOR VAPOR	T12	156.4	155.6
COOLING WATER INLET	T29	49.7	48.5
COOLING WATER OUTLET	T1	93.8	91.1
CONDENSER CONDENSATE	T11	143.2	137.3
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.94	3.70
SECOND EFFECT PRESSURE		-13.72	-13.72
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		5.34	4.79
TOTAL COMPONENT BALANCE		6.18	5.69

Table A-11 Steady State Data for Run OL19 (-10% step in SIFB)

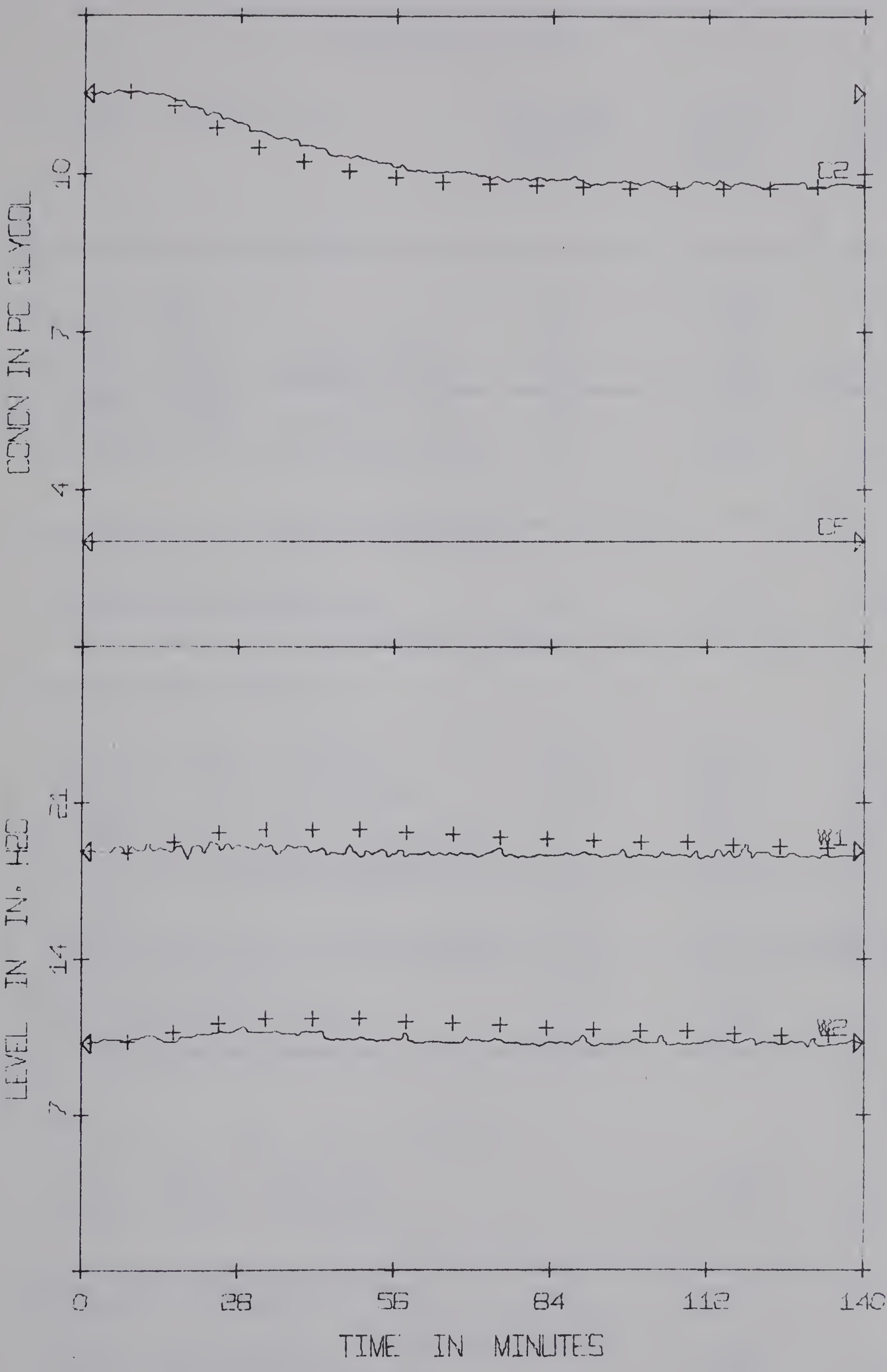


Figure A-10a Transient Data for Run OL19 (-10% step in SIFB)

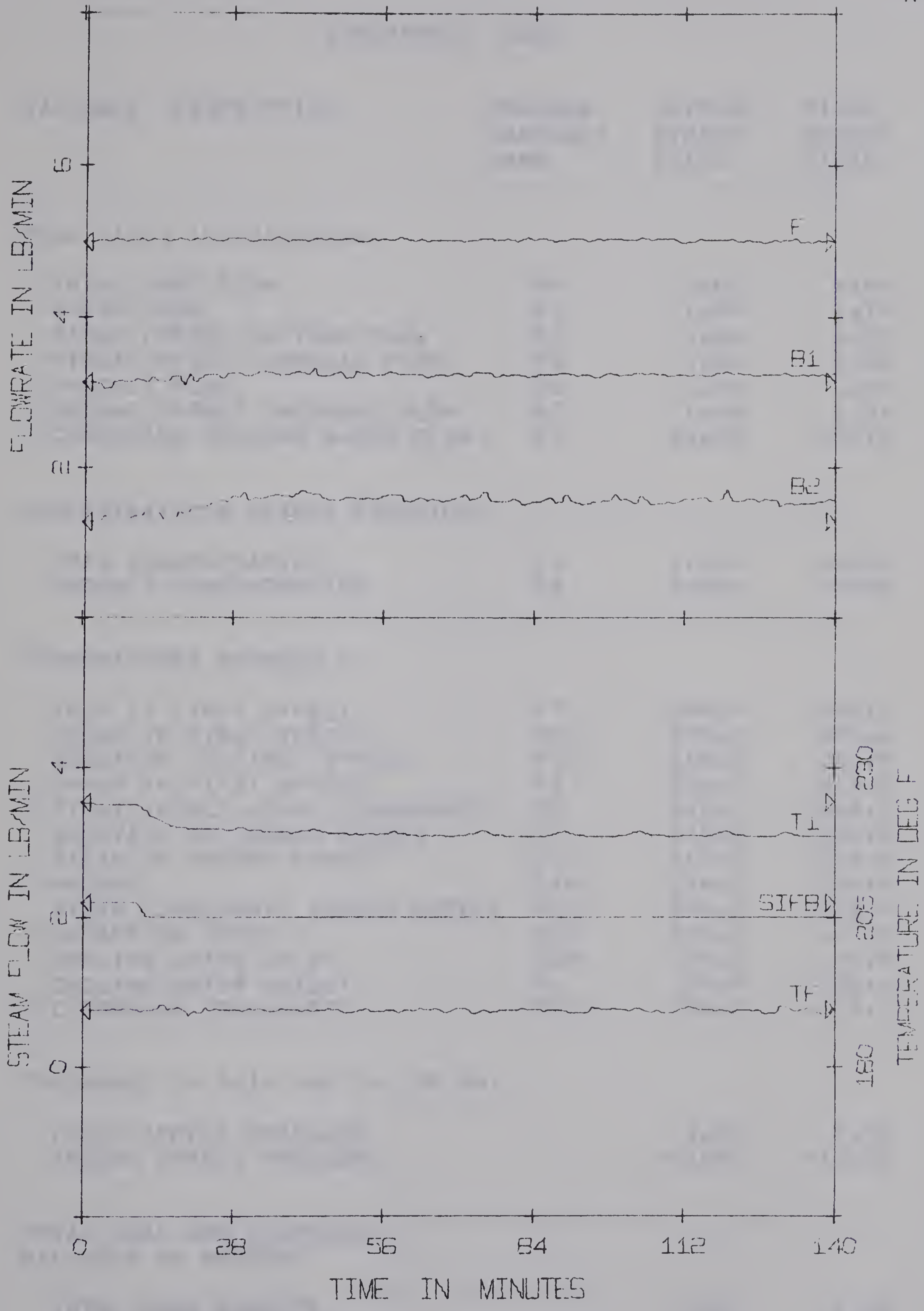


Figure A-10b Transient Data for Run OL19 (-10% step in SIFB)

EXPERIMENT DDC1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.00	4.50
STEAM FLOW	F1	1.94	1.76
FIRST EFFECT BOTTOMS FLOW	F2	3.26	2.91
FIRST EFFECT OVERHEAD FLOW	F5	1.66	1.52
PRODUCT FLOW	F6	1.49	1.44
SECOND EFFECT OVERHEAD FLOW	F7	1.49	1.33
CONDENSER COOLING WATER FLOW	F9	40.08	40.13
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.095	0.095
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.0	189.0
STEAM TO FIRST EFFECT	T15	279.0	280.0
SOLUTION IN FIRST EFFECT	T19	218.0	215.8
VAPOR IN FIRST EFFECT	T2	216.7	214.2
FIRST EFFECT STEAM CONDENSATE	T5	242.1	240.2
SOLUTION TO SECOND EFFECT	T4	178.4	178.8
STEAM TO SECOND EFFECT	T10	217.2	214.8
PRODUCT	T34	154.2	154.2
STEAM CONDENSATE SECOND EFFECT	T28	191.5	189.3
SEPARATOR VAPOR	T12	156.3	157.0
COOLING WATER INLET	T29	49.0	47.8
COOLING WATER OUTLET	T1	89.9	86.3
CONDENSER CONDENSATE	T11	130.8	112.9
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		3.80	2.58
SECOND EFFECT PRESSURE		-13.02	-13.02
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.57	2.53
TOTAL COMPONENT BALANCE		2.80	-2.54

Table A-12 Steady State Data for Run DDC1 (-10% step in F)

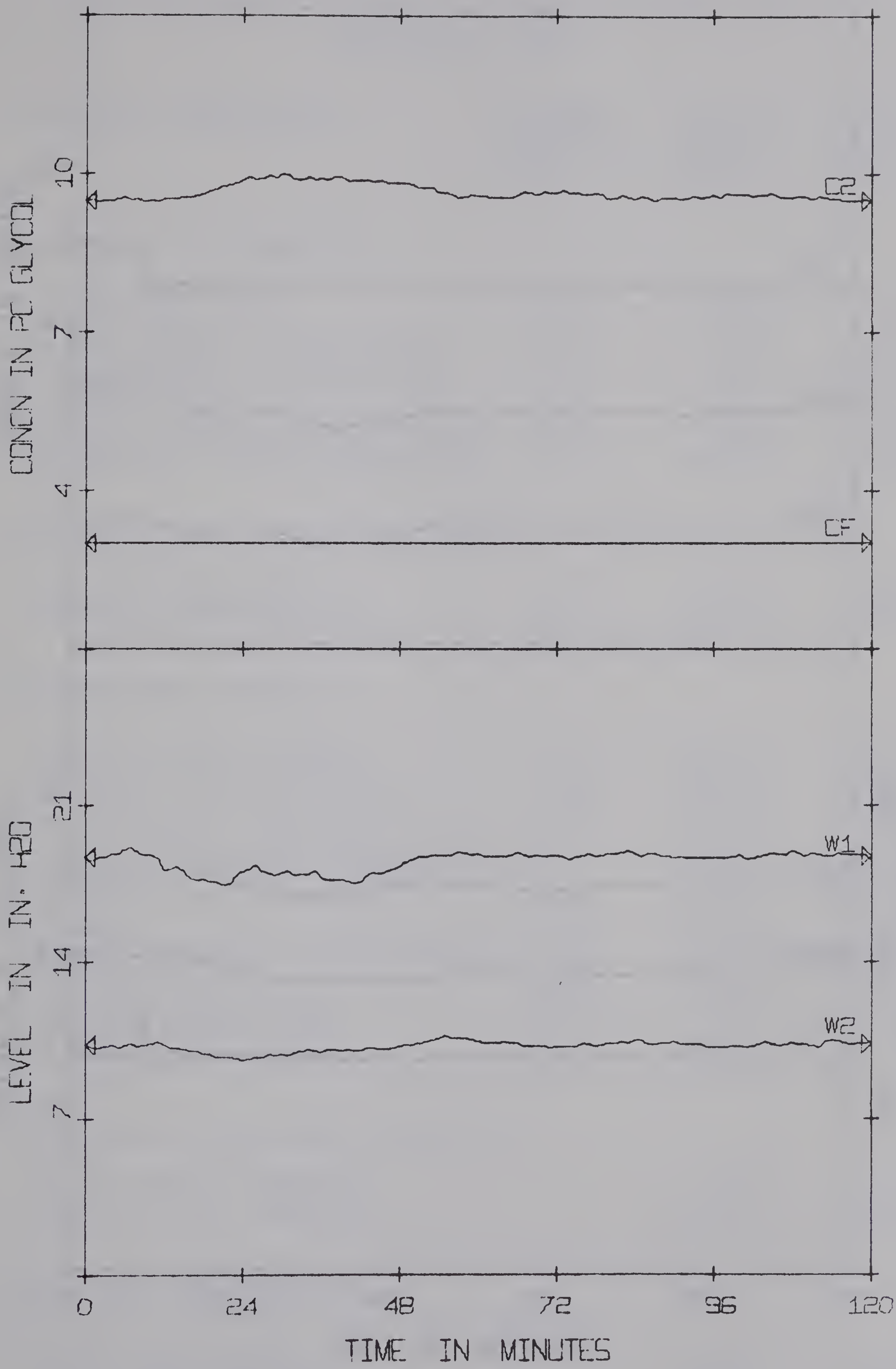


Figure A-11a Transient Data for Run DDC1 (-10% step in F)

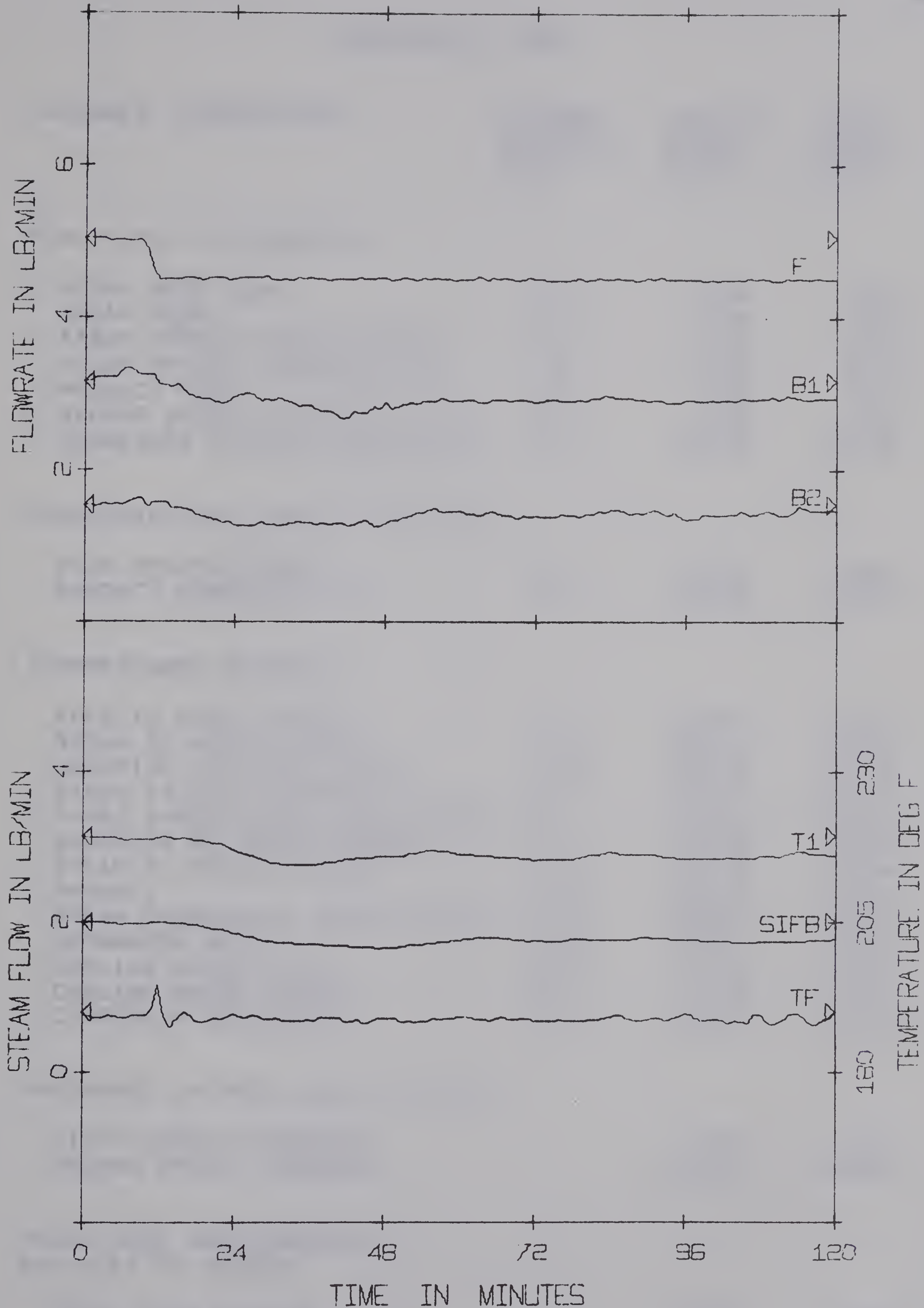


Figure A-11b Transient Data for Run DDC1 (-10% step in F)

EXPERIMENT DDC2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.50	5.01
STEAM FLOW	F1	1.76	1.97
FIRST EFFECT BOTTOMS FLOW	F2	2.91	3.30
FIRST EFFECT OVERHEAD FLOW	F5	1.52	1.70
PRODUCT FLOW	F6	1.44	1.61
SECOND EFFECT OVERHEAD FLOW	F7	1.33	1.47
CONDENSER COOLING WATER FLOW	F9	40.13	39.78
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.095	0.095
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.0	190.3
STEAM TO FIRST EFFECT	T15	280.0	279.4
SOLUTION IN FIRST EFFECT	T19	215.8	217.8
VAPOR IN FIRST EFFECT	T2	214.2	216.4
FIRST EFFECT STEAM CONDENSATE	T5	240.2	241.4
SOLUTION TO SECOND EFFECT	T4	178.8	178.2
STEAM TO SECOND EFFECT	T10	214.8	216.8
PRODUCT	T34	154.2	152.3
STEAM CONDENSATE SECOND EFFECT	T28	189.3	191.2
SEPARATOR VAPOR	T12	157.0	154.1
COOLING WATER INLET	T29	47.8	48.2
COOLING WATER OUTLET	T1	86.3	90.9
CONDENSER CONDENSATE	T11	112.9	137.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.58	3.33
SECOND EFFECT PRESSURE		-13.02	-12.80
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.53	3.21
TOTAL COMPONENT BALANCE		-2.54	0.00

Table A-13 Steady State Data for Run DDC2 (+10% step in F)

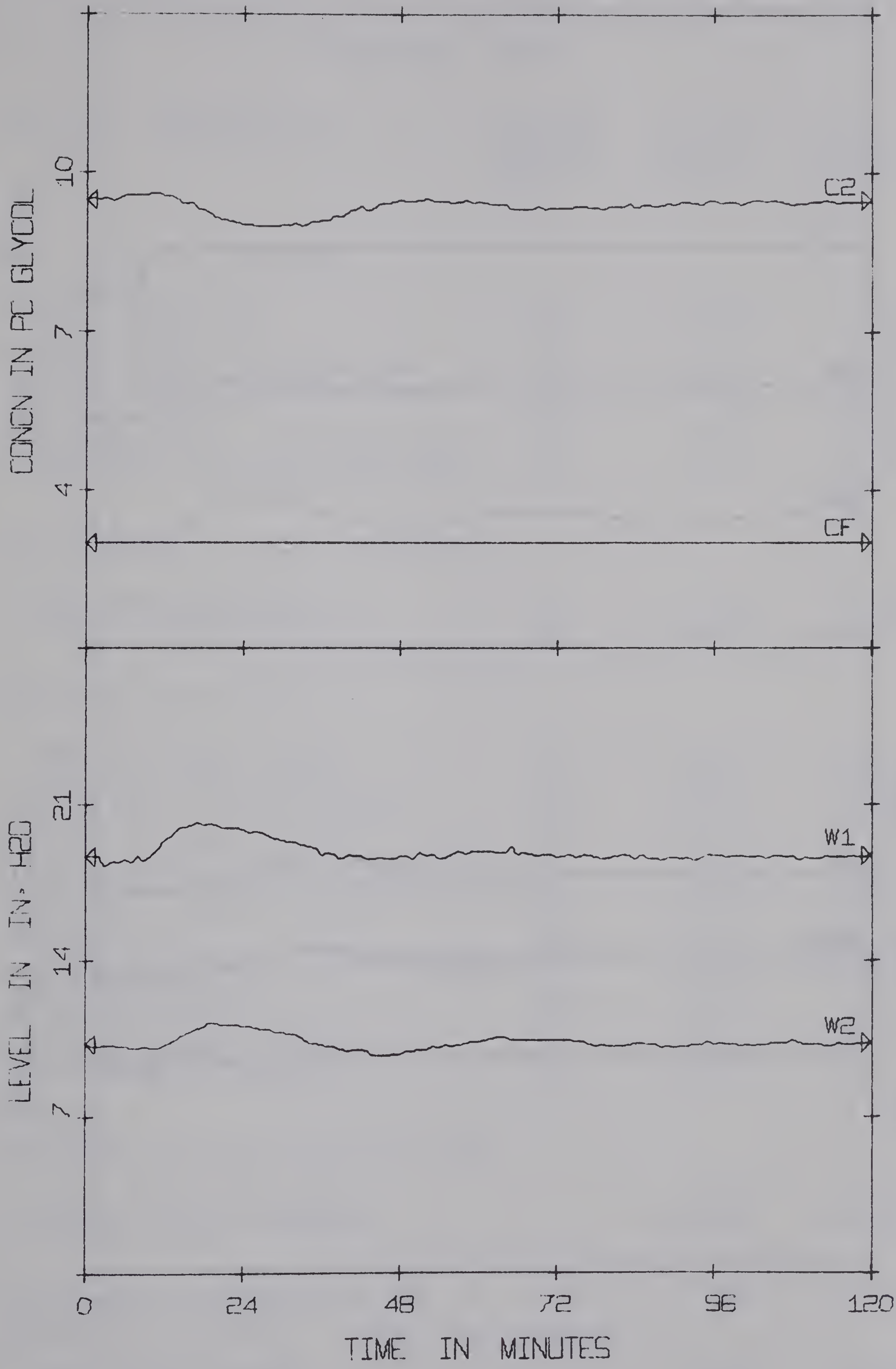
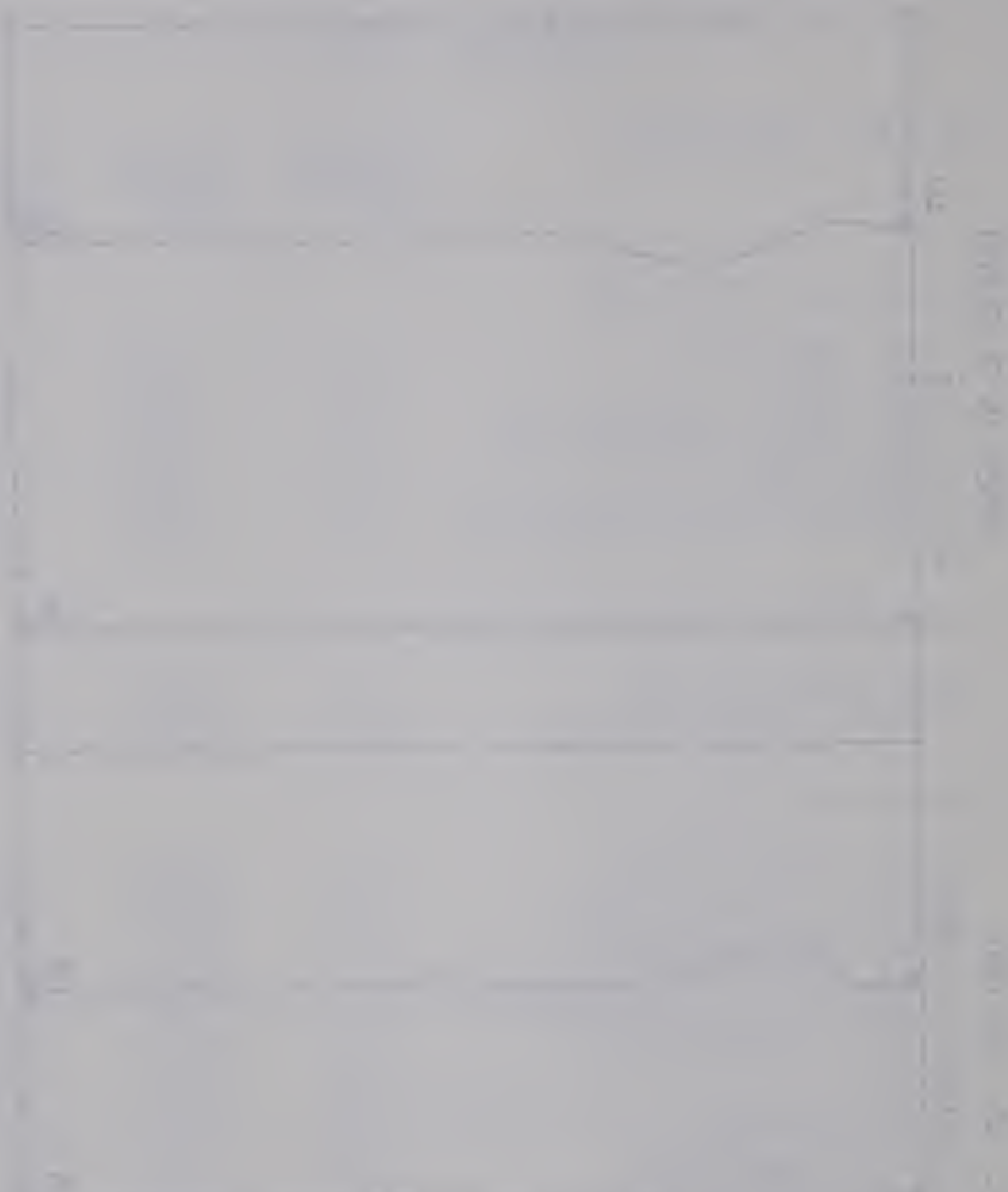


Figure A-12a Transient Data for Run DDC2 (+10% step in F)



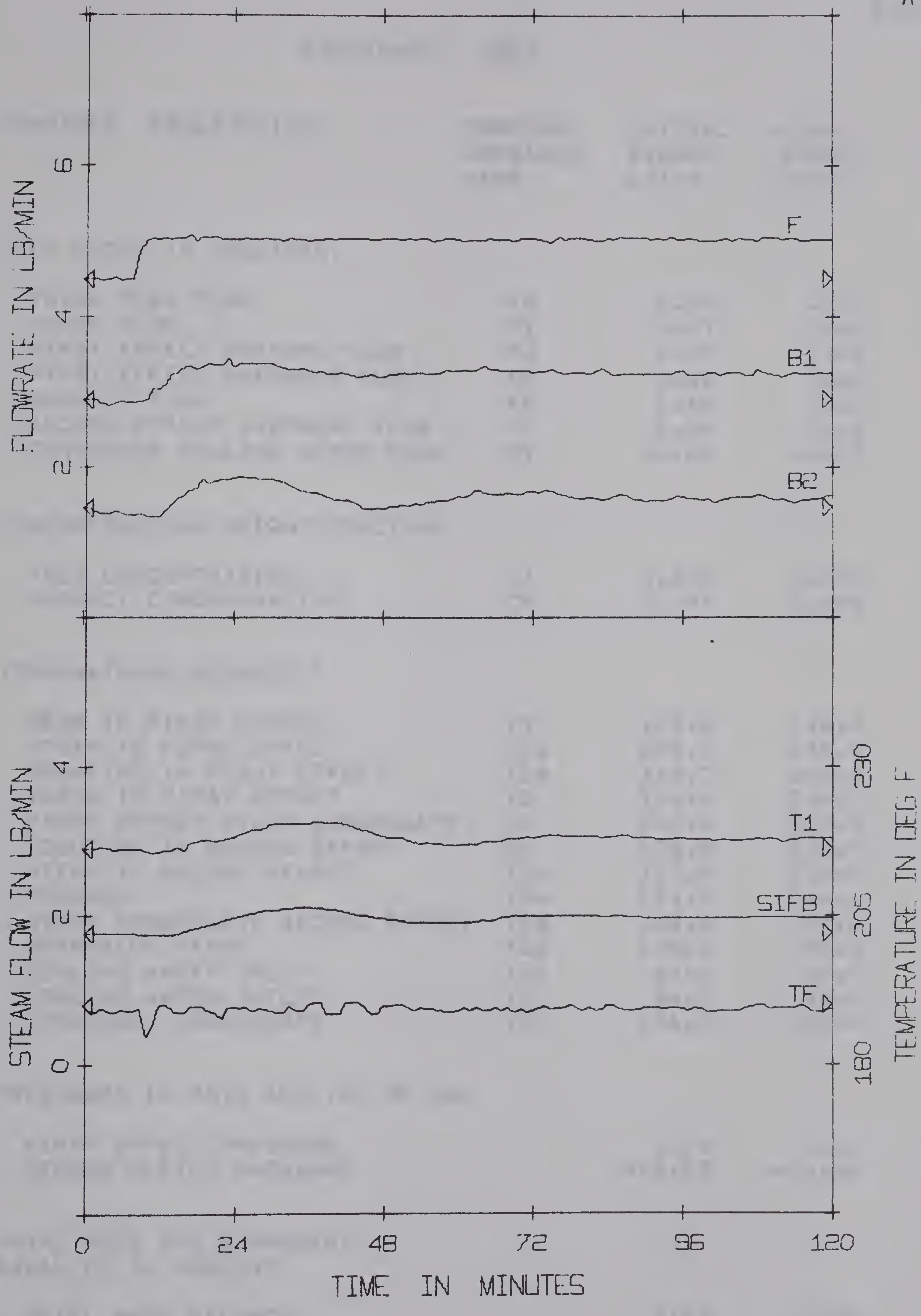
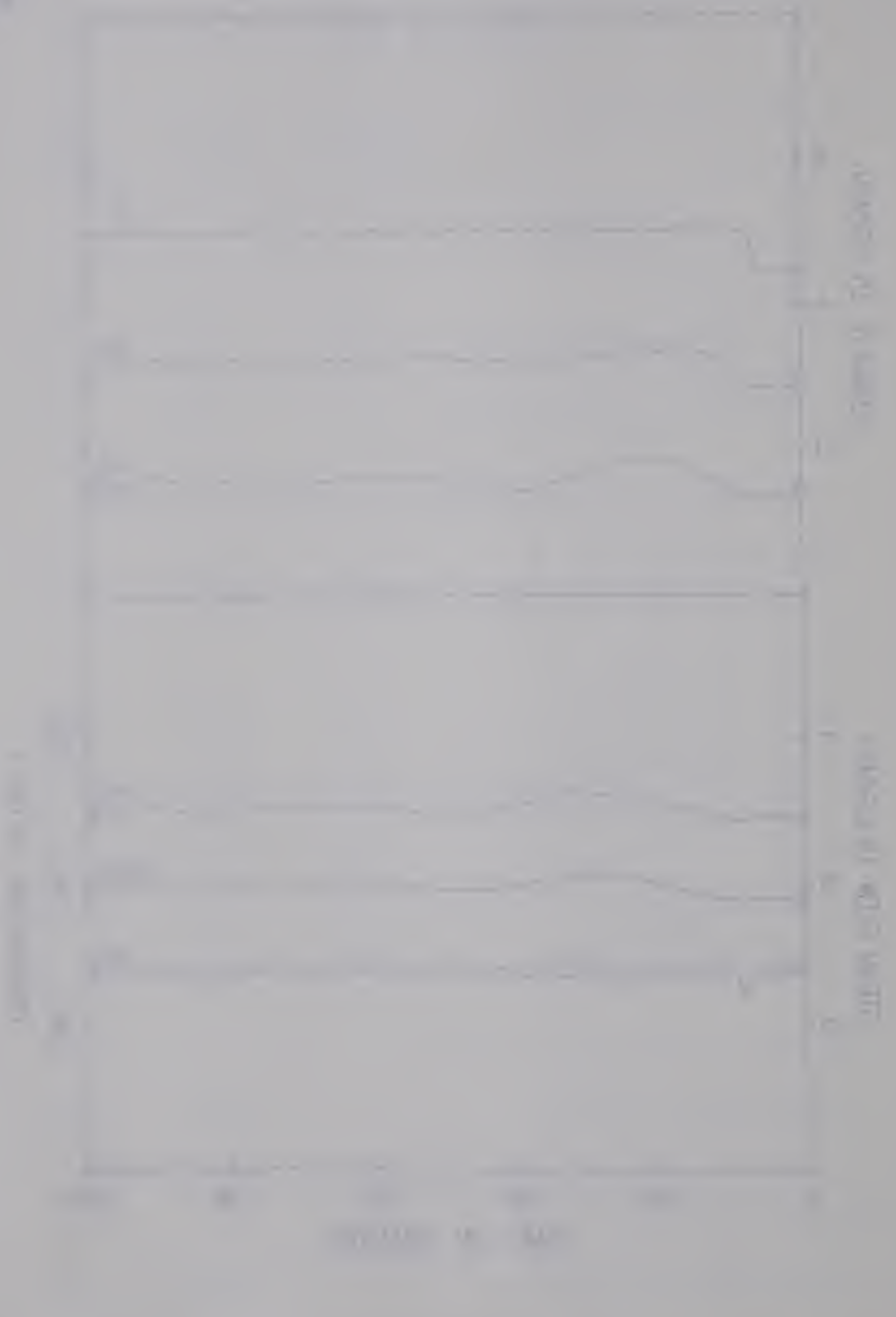


Figure A-12b Transient Data for Run DDC2 (+10% step in F)



EXPERIMENT DDC3

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.04	4.51
STEAM FLOW	F1	1.97	1.69
FIRST EFFECT BOTTOMS FLOW	F2	3.27	2.82
FIRST EFFECT OVERHEAD FLOW	F5	1.68	1.58
PRODUCT FLOW	F6	1.60	1.40
SECOND EFFECT OVERHEAD FLOW	F7	1.54	1.26
CONDENSER COOLING WATER FLOW	F9	40.01	40.13
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.095	0.095
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.3	188.9
STEAM TO FIRST EFFECT	T15	279.4	280.9
SOLUTION IN FIRST EFFECT	T19	218.7	216.0
VAPOR IN FIRST EFFECT	T2	216.9	214.1
FIRST EFFECT STEAM CONDENSATE	T5	241.6	235.8
SOLUTION TO SECOND EFFECT	T4	178.6	178.9
STEAM TO SECOND EFFECT	T10	217.3	214.5
PRODUCT	T34	153.9	154.1
STEAM CONDENSATE SECOND EFFECT	T28	193.8	189.6
SEPARATOR VAPOR	T12	156.1	157.1
COOLING WATER INLET	T29	52.4	50.7
COOLING WATER OUTLET	T1	94.1	87.9
CONDENSER CONDENSATE	T11	136.9	106.4
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		3.75	2.10
SECOND EFFECT PRESSURE		-13.02	-12.98
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.52	4.26
TOTAL COMPONENT BALANCE		-1.07	-3.77

Table A-14 Steady State Data for Run DDC3 (-10% step in F)

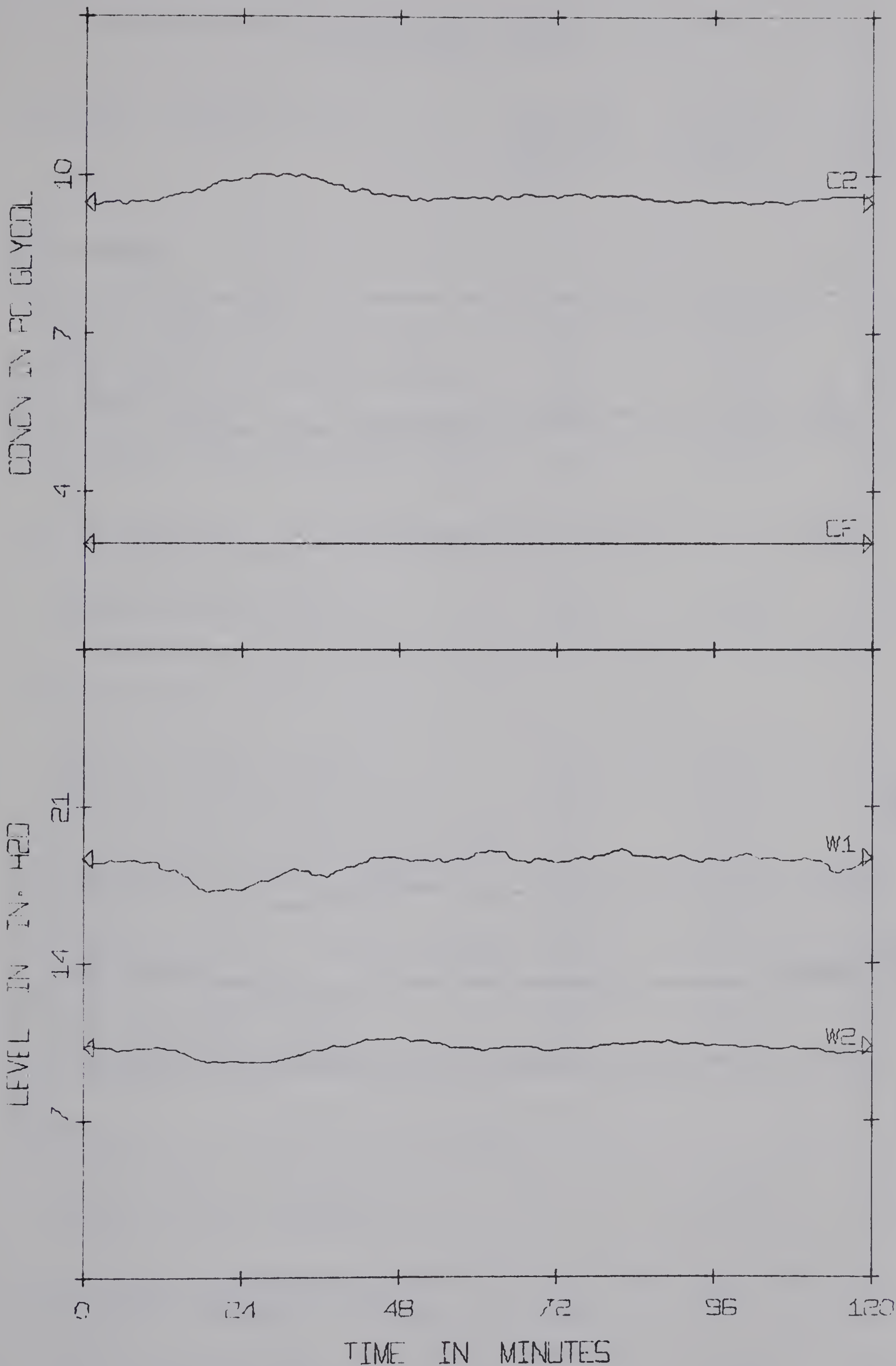


Figure A-13a Transient Data for Run DDC3 (-10% step in F)

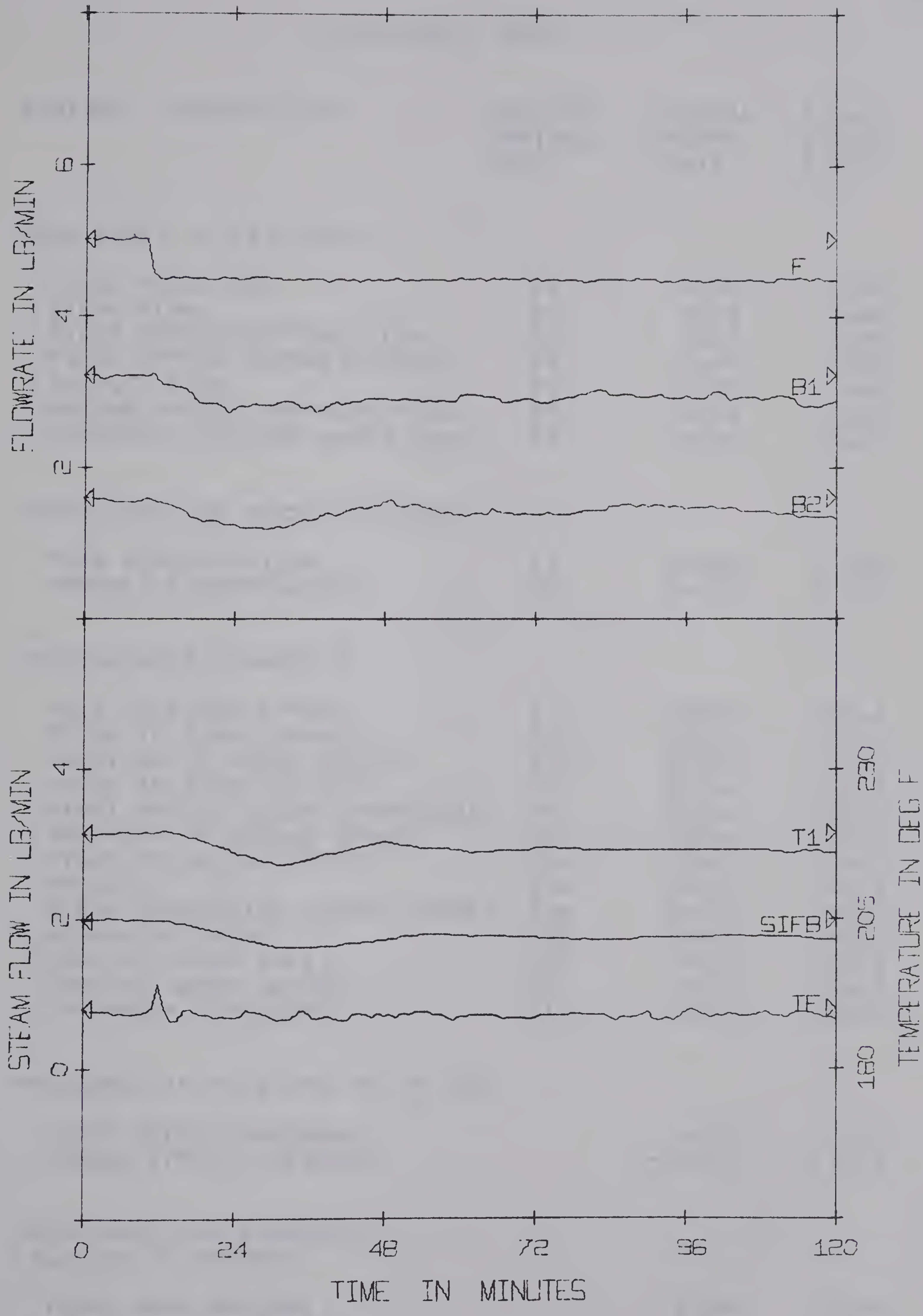


Figure A-13b Transient Data for Run DDC3 (-10% step in F)

EXPERIMENT DDC4

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	4.48
STEAM FLOW	F1	2.19	1.66
FIRST EFFECT BOTTOMS FLOW	F2	3.57	2.95
FIRST EFFECT OVERHEAD FLOW	F5	1.87	1.51
PRODUCT FLOW	F6	1.70	1.48
SECOND EFFECT OVERHEAD FLOW	F7	1.72	1.33
CONDENSER COOLING WATER FLOW	F9	39.50	40.03
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.095	0.095
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.9	188.6
STEAM TO FIRST EFFECT	T15	300.4	305.1
SOLUTION IN FIRST EFFECT	T19	227.8	217.6
VAPOR IN FIRST EFFECT	T2	225.6	215.2
FIRST EFFECT STEAM CONDENSATE	T5	252.8	236.9
SOLUTION TO SECOND EFFECT	T4	182.6	181.1
STEAM TO SECOND EFFECT	T10	226.2	215.5
PRODUCT	T34	158.3	156.8
STEAM CONDENSATE SECOND EFFECT	T28	201.3	191.1
SEPARATOR VAPOR	T12	160.3	160.0
COOLING WATER INLET	T29	50.5	52.8
COOLING WATER OUTLET	T1	97.5	90.7
CONDENSER CONDENSATE	T11	151.0	102.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		9.25	2.62
SECOND EFFECT PRESSURE		-14.97	-15.03
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.39	2.66
TOTAL COMPONENT BALANCE		1.85	-6.38

Table A-15 Steady State Data for Run DDC4 (-20% step in F)

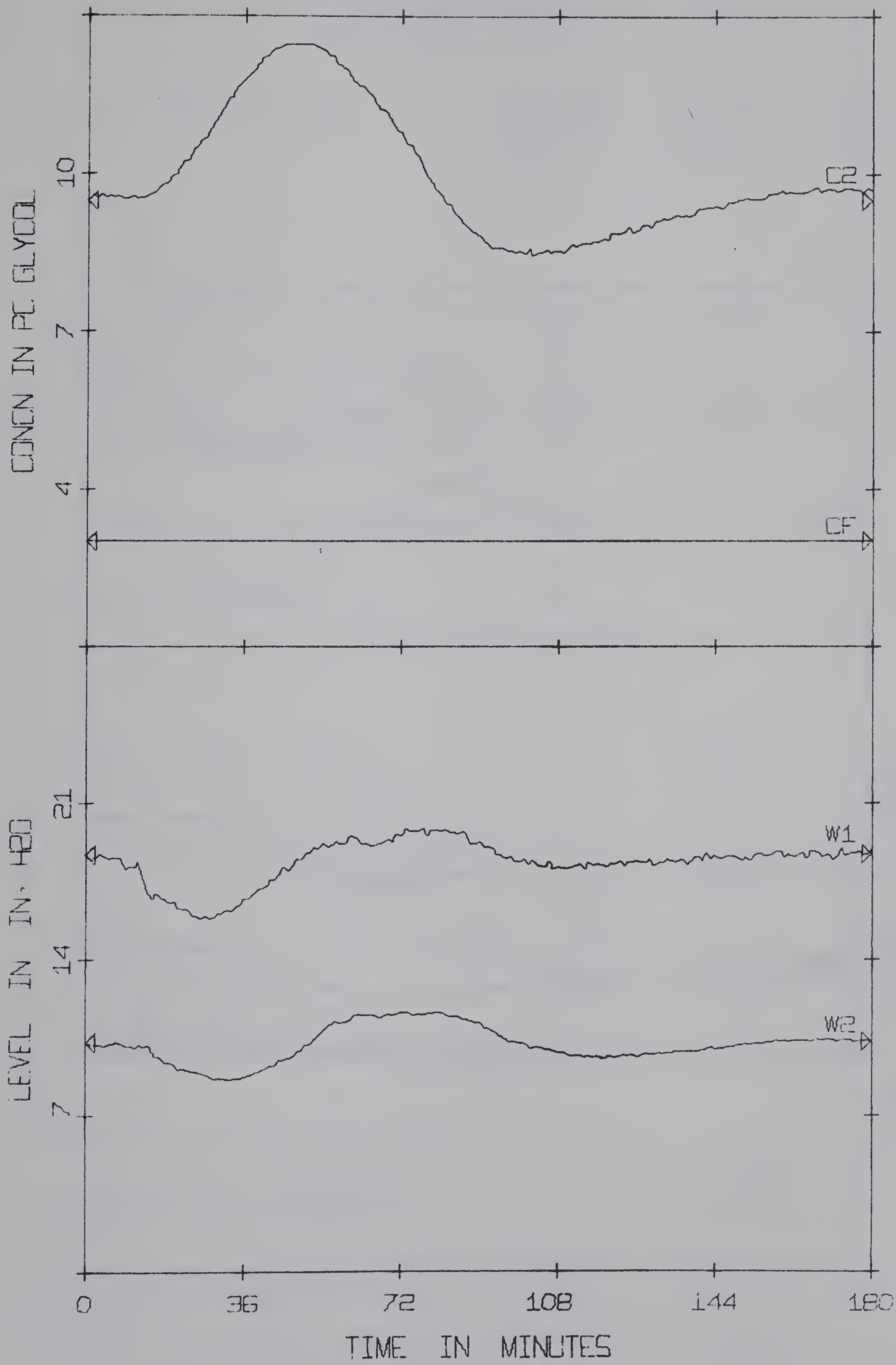
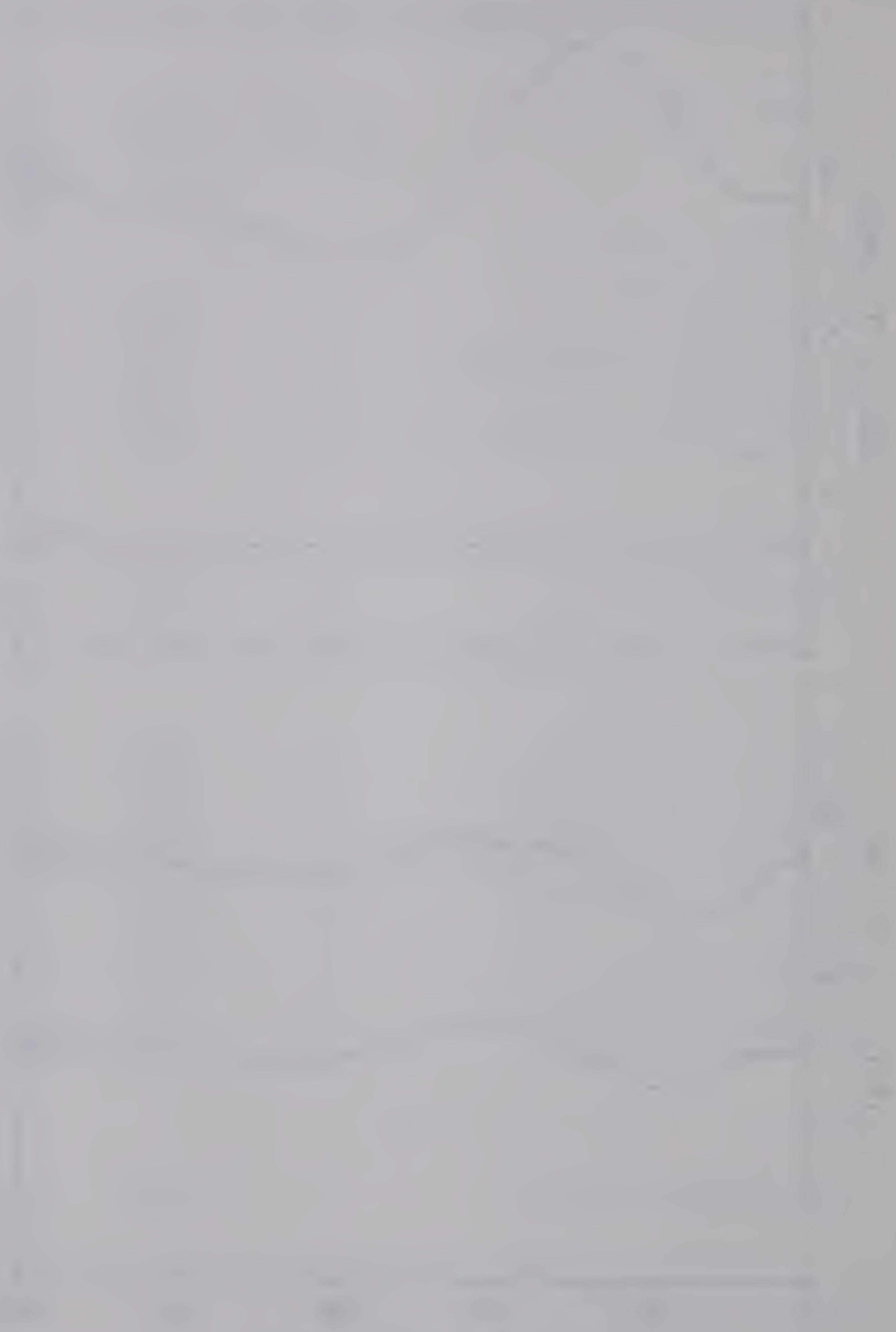


Figure A-14a Transient Data for Run DDC4 (-20% step in F)



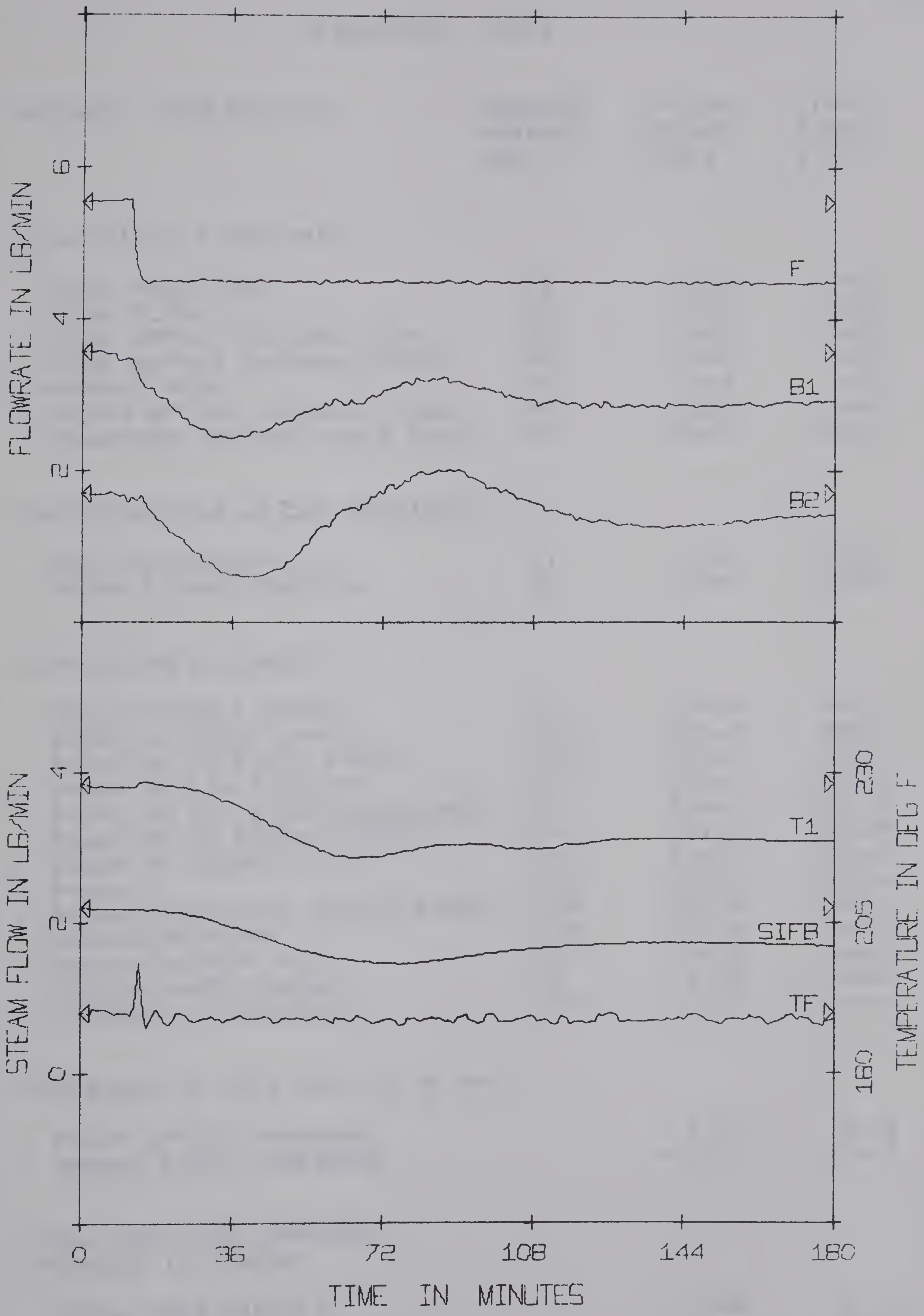


Figure A-14b Transient Data for Run DDC4 (-20% step in F)

EXPERIMENT DDC5

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.47	5.54
STEAM FLOW	F1	1.73	2.25
FIRST EFFECT BOTTOMS FLOW	F2	2.92	3.66
FIRST EFFECT OVERHEAD FLOW	F5	1.49	1.81
PRODUCT FLOW	F6	1.48	1.76
SECOND EFFECT OVERHEAD FLOW	F7	1.27	1.66
CONDENSER COOLING WATER FLOW	F9	40.15	40.37
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.4	189.0
STEAM TO FIRST EFFECT	T15	304.7	299.7
SOLUTION IN FIRST EFFECT	T19	217.2	226.0
VAPOR IN FIRST EFFECT	T2	215.7	224.2
FIRST EFFECT STEAM CONDENSATE	T5	238.0	251.2
SOLUTION TO SECOND EFFECT	T4	181.4	182.8
STEAM TO SECOND EFFECT	T10	216.2	224.9
PRODUCT	T34	157.7	158.4
STEAM CONDENSATE SECOND EFFECT	T28	191.6	200.0
SEPARATOR VAPOR	T12	160.6	160.5
COOLING WATER INLET	T29	49.9	50.7
COOLING WATER OUTLET	T1	86.9	96.0
CONDENSER CONDENSATE	T11	113.0	146.4
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.73	8.18
SECOND EFFECT PRESSURE		-15.01	-15.03
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.14	3.59
TOTAL COMPONENT BALANCE		1.24	0.57

Table A-16   Steady State Data for Run DDC5 (+20% step in F)

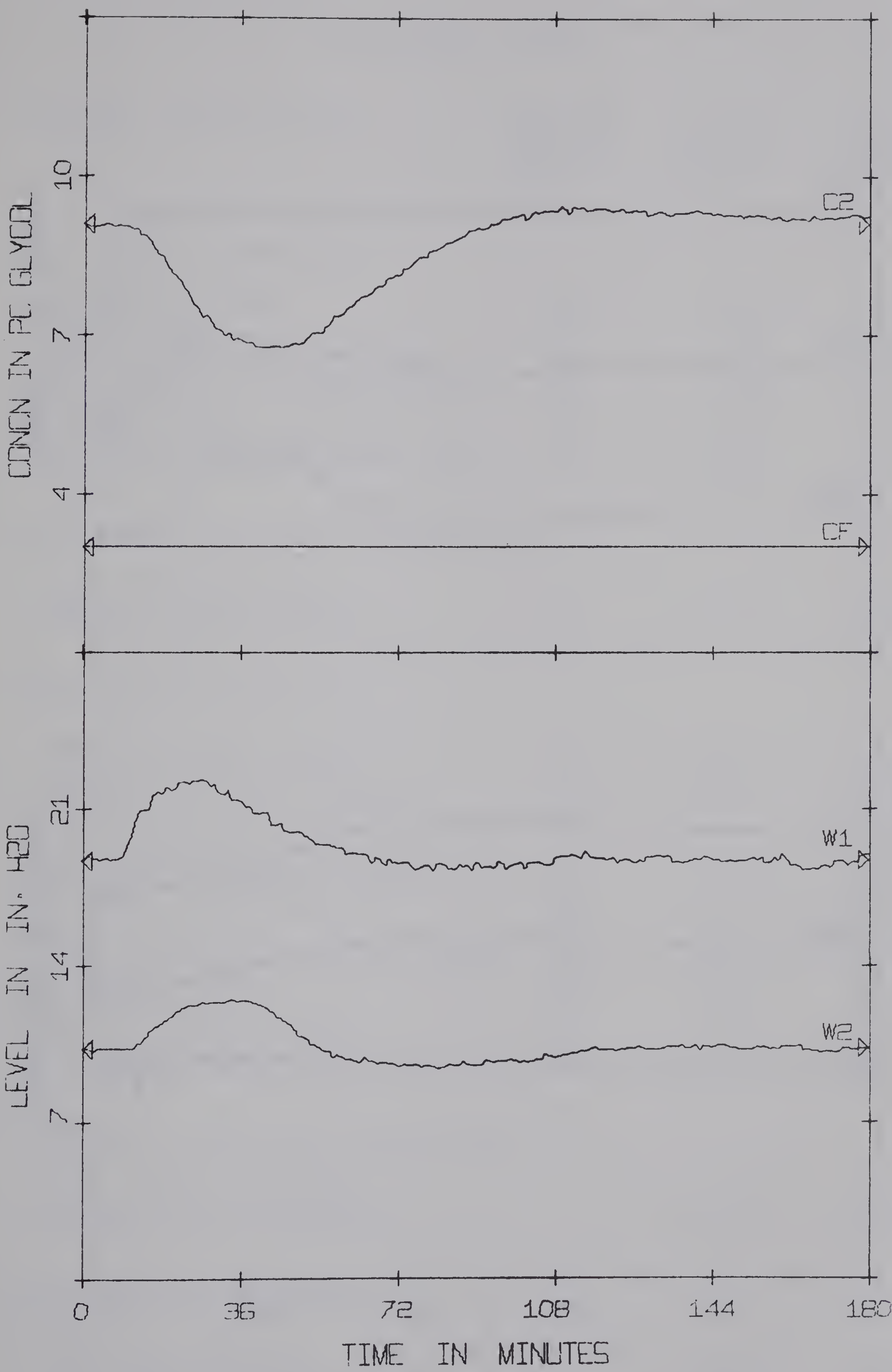


Figure A-15a Transient Data for Run DDC5 (+20% step in F)

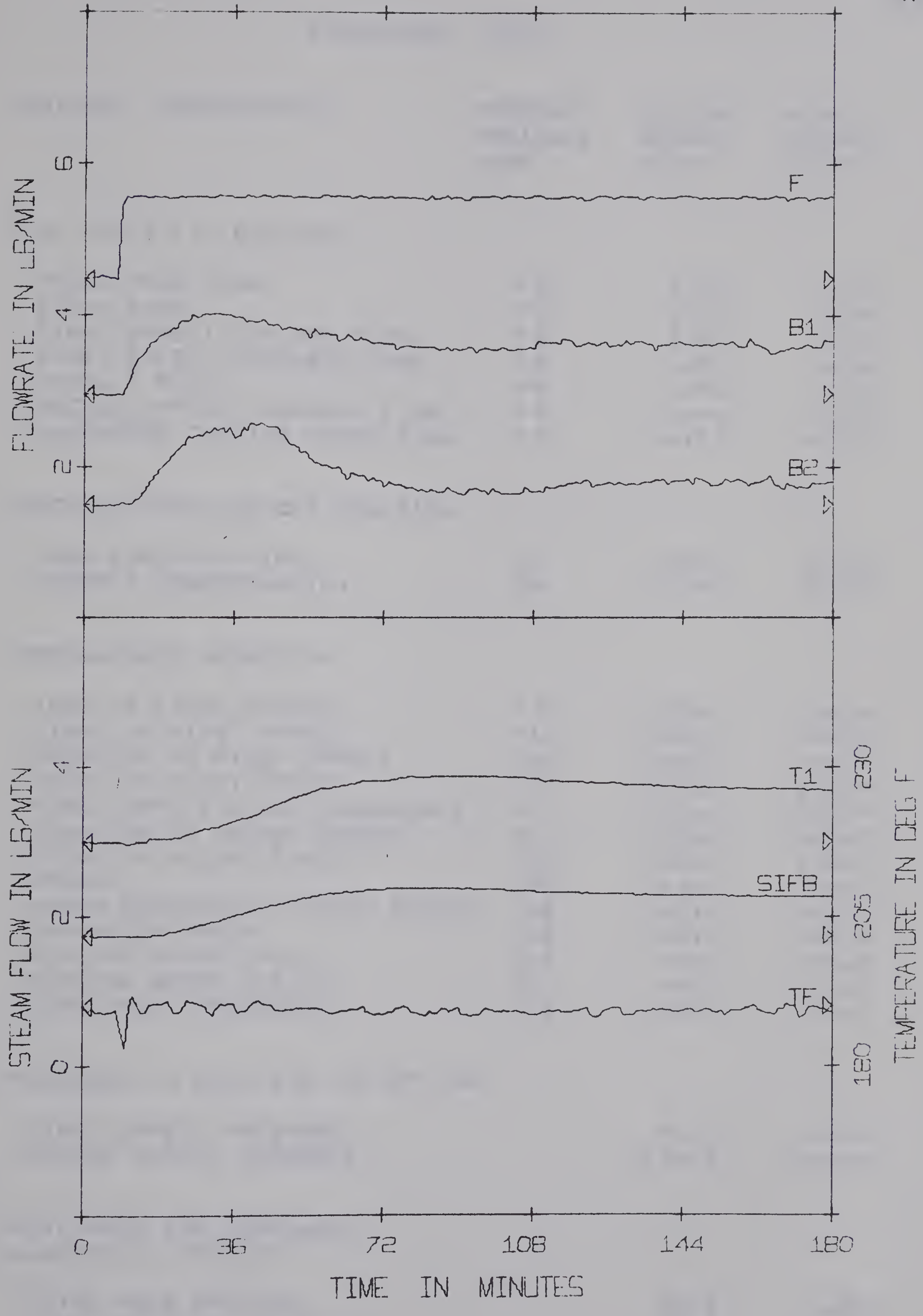


Figure A-15b Transient Data for Run DDC5 (+20% step in F)

EXPERIMENT DDC6

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	4.46
STEAM FLOW	F1	2.25	1.74
FIRST EFFECT BOTTOMS FLOW	F2	3.66	3.02
FIRST EFFECT OVERHEAD FLOW	F5	1.81	1.50
PRODUCT FLOW	F6	1.76	1.52
SECOND EFFECT OVERHEAD FLOW	F7	1.66	1.27
CONDENSER COOLING WATER FLOW	F9	40.37	40.15
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.0	188.6
STEAM TO FIRST EFFECT	T15	299.7	304.5
SOLUTION IN FIRST EFFECT	T19	226.0	217.6
VAPOR IN FIRST EFFECT	T2	224.2	215.9
FIRST EFFECT STEAM CONDENSATE	T5	251.2	237.9
SOLUTION TO SECOND EFFECT	T4	182.8	181.7
STEAM TO SECOND EFFECT	T10	224.9	216.4
PRODUCT	T34	158.4	158.1
STEAM CONDENSATE SECOND EFFECT	T28	200.0	191.7
SEPARATOR VAPOR	T12	160.5	160.9
COOLING WATER INLET	T29	50.7	49.3
COOLING WATER OUTLET	T1	96.0	87.0
CONDENSER CONDENSATE	T11	146.4	113.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		8.18	2.84
SECOND EFFECT PRESSURE		-15.03	-14.96
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.59	1.98
TOTAL COMPONENT BALANCE		0.57	-1.77

Table A-17 Steady State Data for Run DDC6 (-20% step in F)

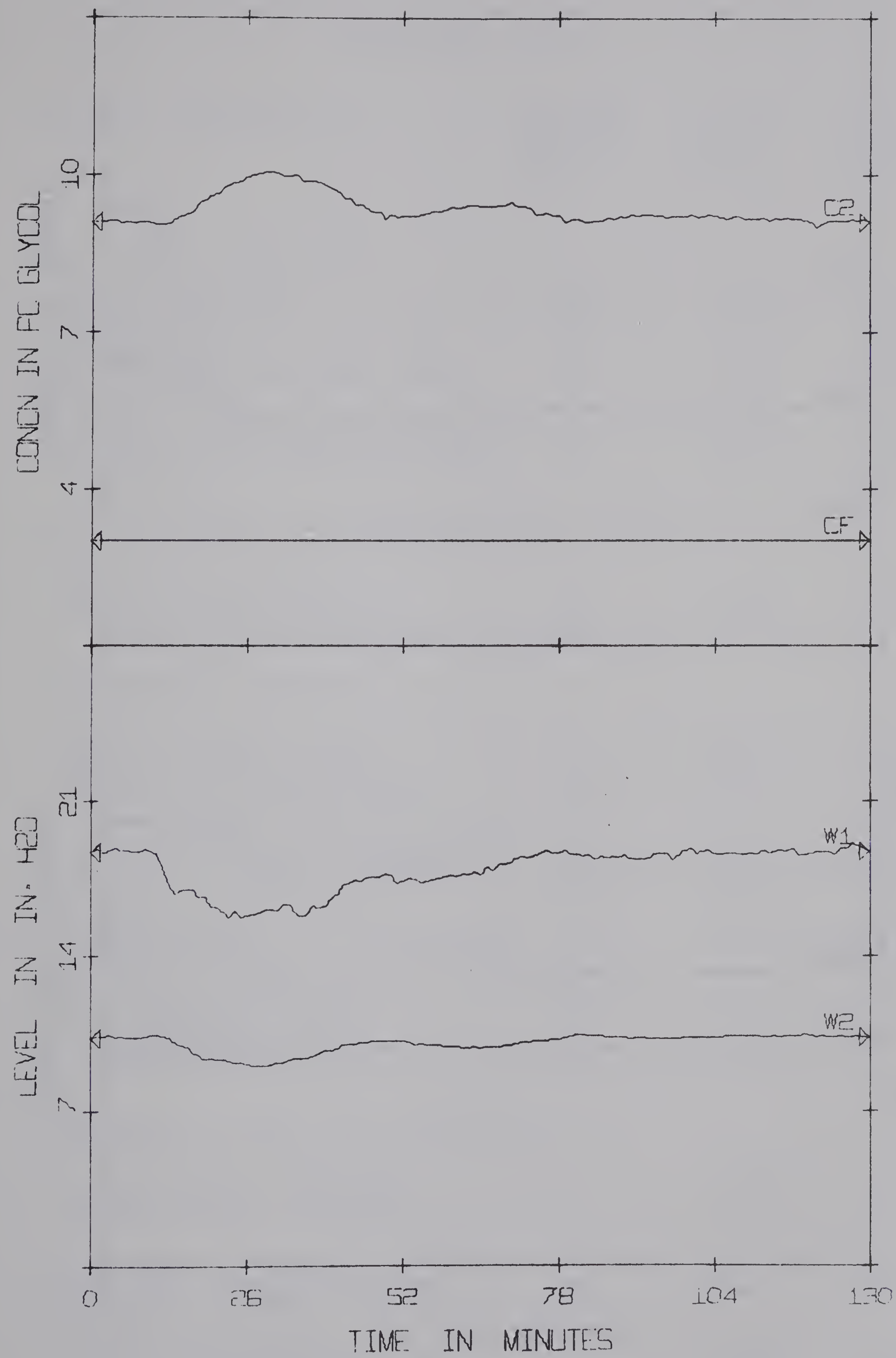


Figure A-16a Transient Data for Run DDC6 (-20% step in F)

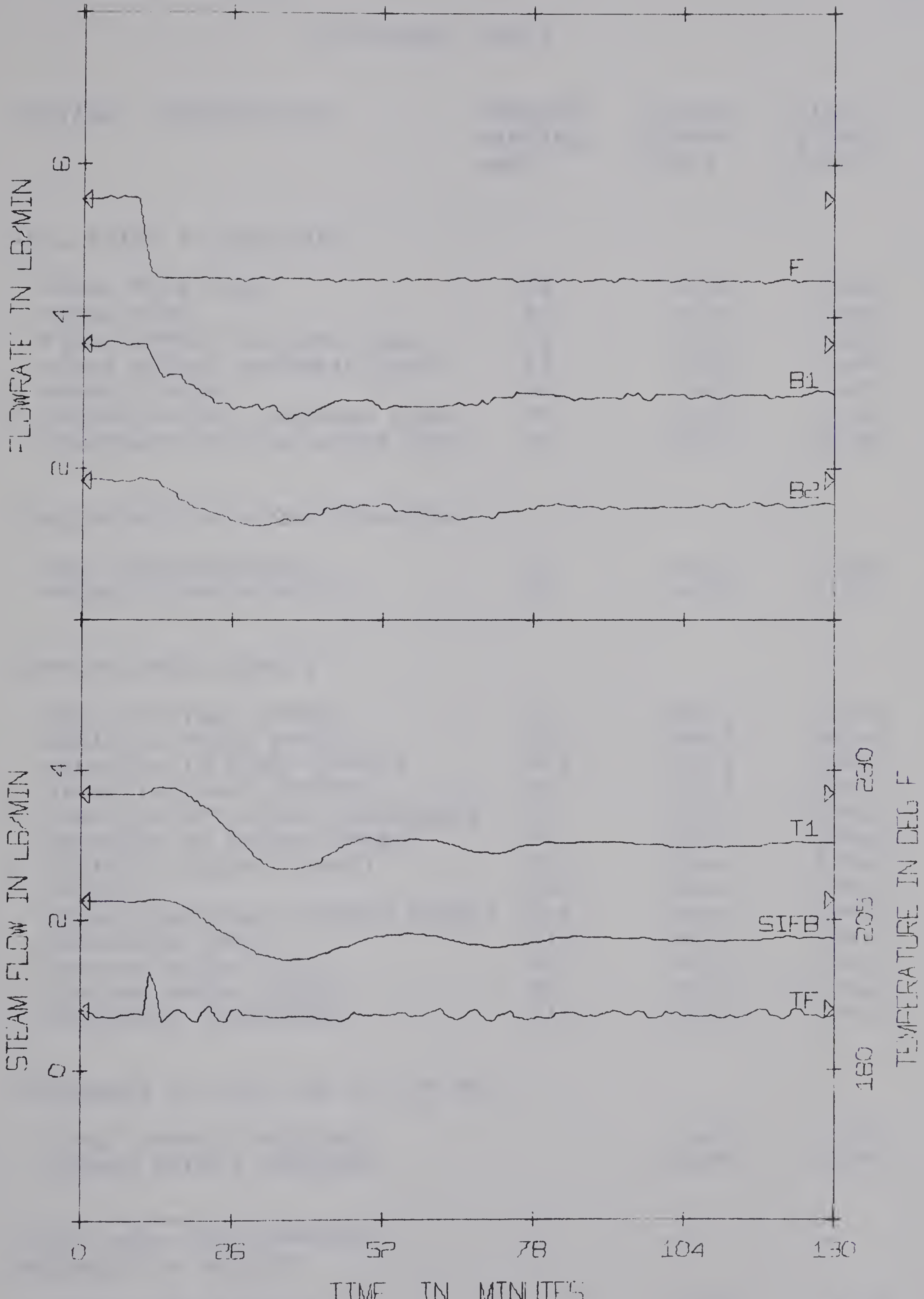


Figure A-16b Transient Data for Run DDC6 (-20% step in F)

EXPERIMENT DDC7

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.46	5.56
STEAM FLOW	F1	1.74	2.21
FIRST EFFECT BOTTOMS FLOW	F2	3.02	3.63
FIRST EFFECT OVERHEAD FLOW	F5	1.50	1.79
PRODUCT FLOW	F6	1.52	1.79
SECOND EFFECT OVERHEAD FLOW	F7	1.27	1.61
CONDENSER COOLING WATER FLOW	F9	40.15	40.35
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.6	189.3
STEAM TO FIRST EFFECT	T15	304.5	300.2
SOLUTION IN FIRST EFFECT	T19	217.6	225.8
VAPOR IN FIRST EFFECT	T2	215.9	224.0
FIRST EFFECT STEAM CONDENSATE	T5	237.9	250.6
SOLUTION TO SECOND EFFECT	T4	181.7	183.0
STEAM TO SECOND EFFECT	T10	216.4	224.6
PRODUCT	T34	158.1	159.1
STEAM CONDENSATE SECOND EFFECT	T28	191.7	199.3
SEPARATOR VAPOR	T12	160.9	160.7
COOLING WATER INLET	T29	49.3	49.0
COOLING WATER OUTLET	T1	87.0	94.9
CONDENSER CONDENSATE	T11	113.0	146.5
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.84	7.94
SECOND EFFECT PRESSURE		-14.96	-14.95
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		1.98	3.35
TOTAL COMPONENT BALANCE		-1.77	-0.17

Table A-18 Steady State Data for Run DDC7 (+20% step in F)

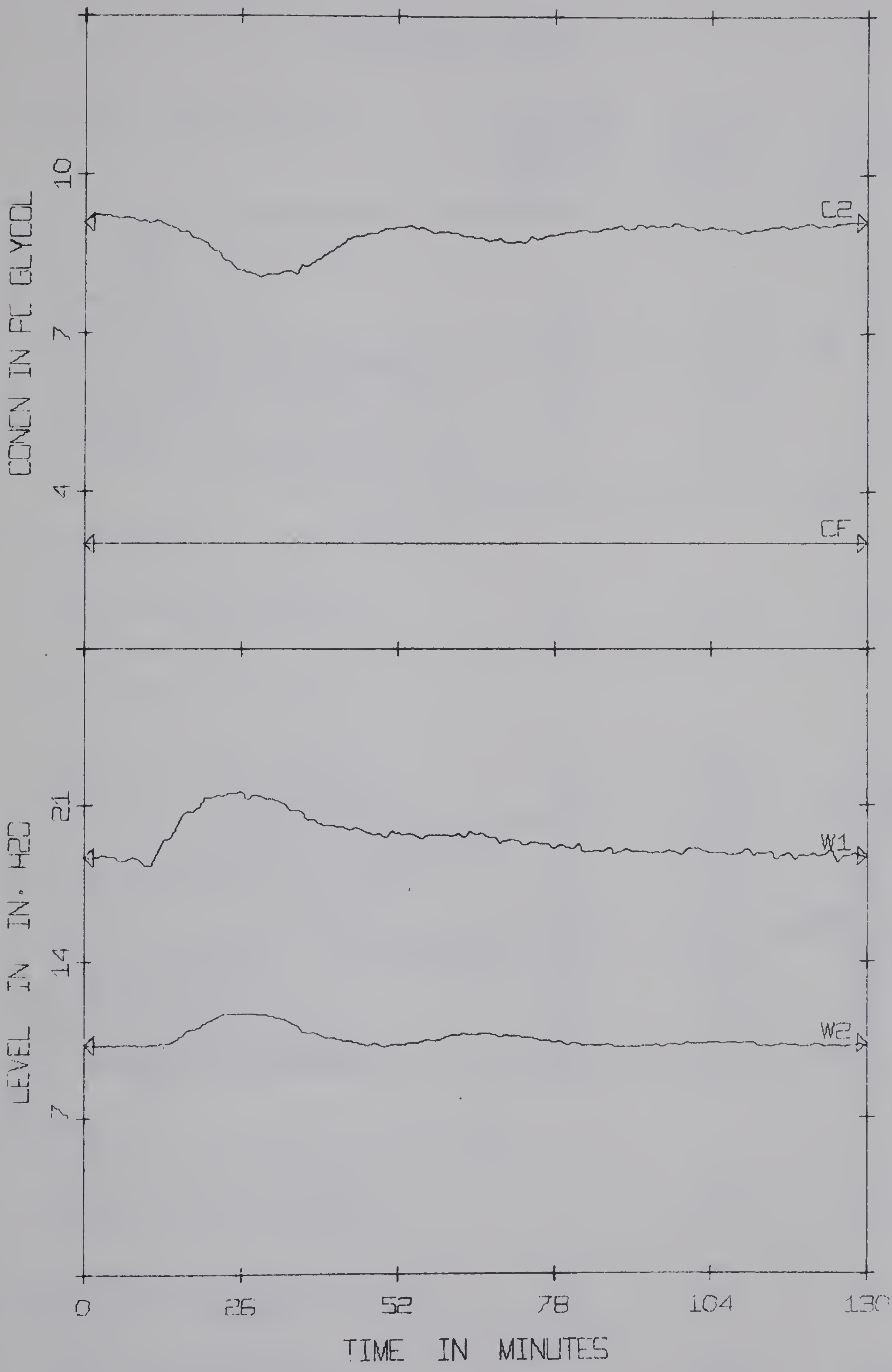


Figure A-17a Transient Data for Run DDC7 (+20% step in F)

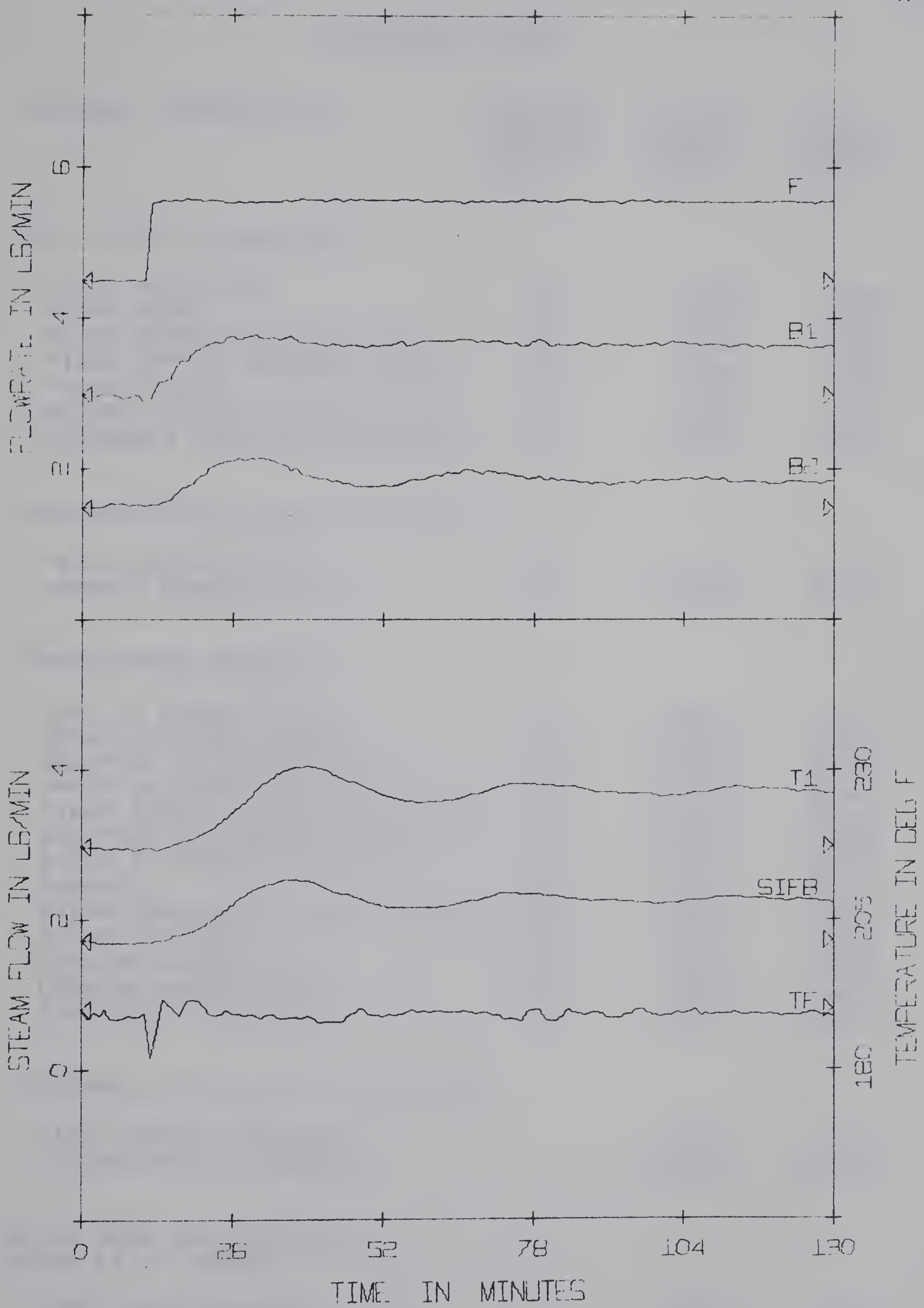


Figure A-17b Transient Data for Run DDC7 (+20% step in F)

EXPERIMENT DDC8

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.52	5.53
STEAM FLOW	F1	1.77	2.23
FIRST EFFECT BOTTOMS FLOW	F2	2.95	3.63
FIRST EFFECT OVERHEAD FLOW	F5	1.55	1.81
PRODUCT FLOW	F6	1.46	1.77
SECOND EFFECT OVERHEAD FLOW	F7	1.36	1.68
CONDENSER COOLING WATER FLOW	F9	39.53	40.13
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.8	189.5
STEAM TO FIRST EFFECT	T15	304.9	299.4
SOLUTION IN FIRST EFFECT	T19	217.4	226.3
VAPOR IN FIRST EFFECT	T2	215.5	224.2
FIRST EFFECT STEAM CONDENSATE	T5	237.9	251.0
SOLUTION TO SECOND EFFECT	T4	180.7	183.2
STEAM TO SECOND EFFECT	T10	216.1	224.8
PRODUCT	T34	156.1	158.4
STEAM CONDENSATE SECOND EFFECT	T28	189.8	199.7
SEPARATOR VAPOR	T12	159.1	160.2
COOLING WATER INLET	T29	53.9	51.2
COOLING WATER OUTLET	T1	93.5	94.6
CONDENSER CONDENSATE	T11	124.7	143.7
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.36	7.44
SECOND EFFECT PRESSURE		-15.04	-15.02
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.04	3.78
TOTAL COMPONENT BALANCE		1.63	1.37

Table A-19 Steady State Data for Run DDC8 (+20% step in F)

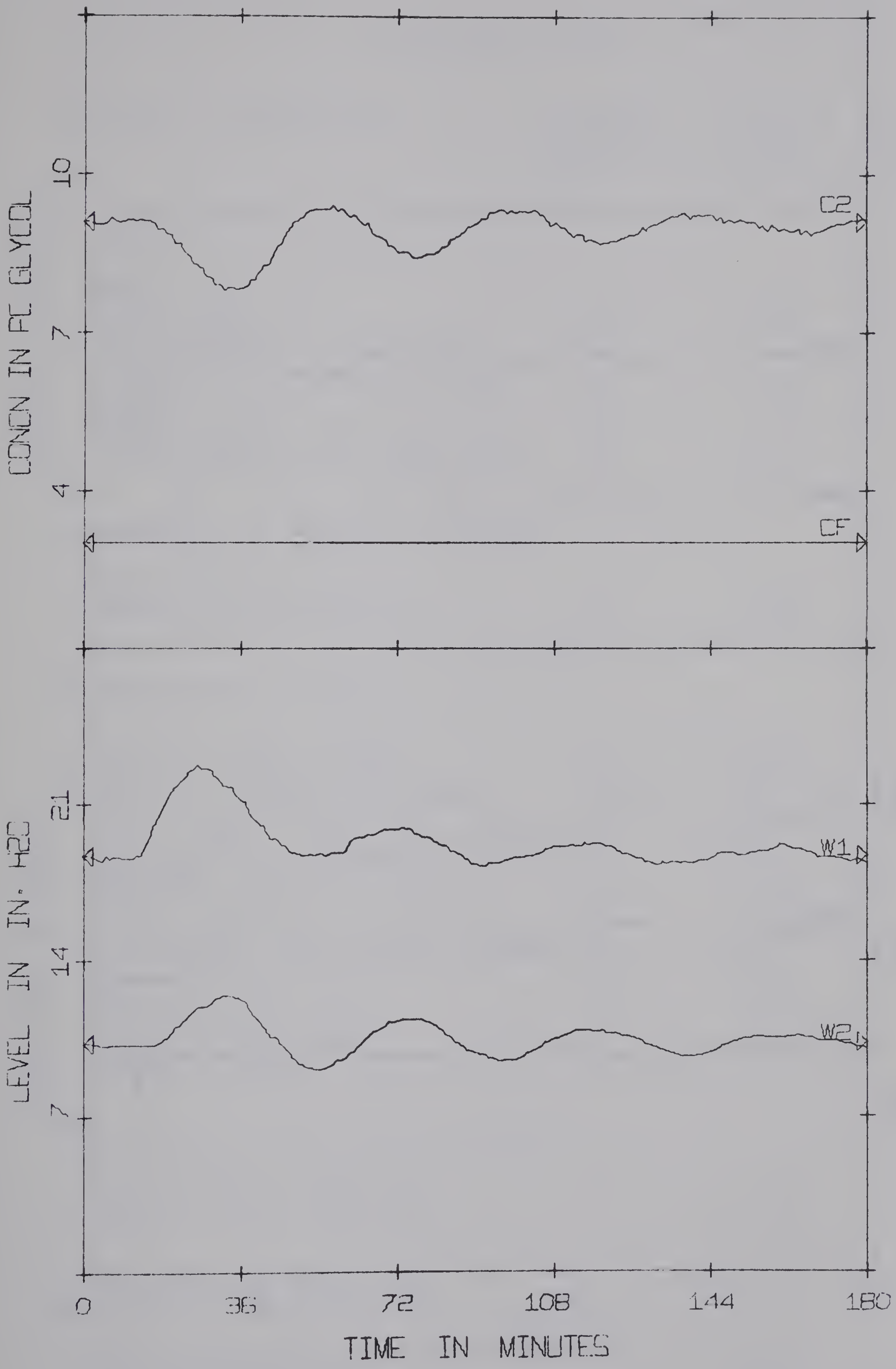


Figure A-18a Transient Data for Run DDC8 (+20% step in F)

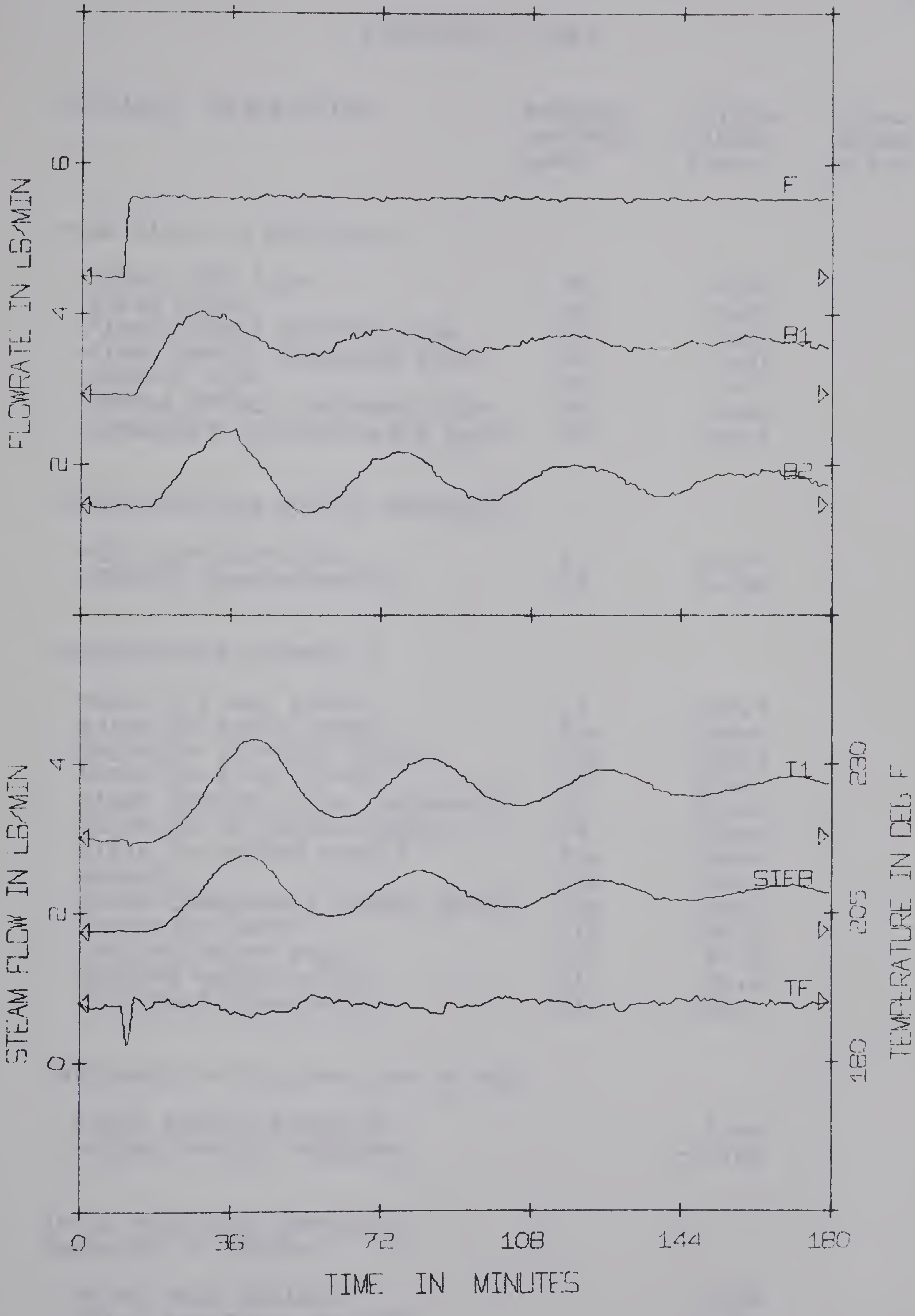


Figure A-18b Transient Data for Run DDC8 (+20% step in F)

EXPERIMENT DDC9

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.53	
STEAM FLOW	F1	2.23	
FIRST EFFECT BOTTOMS FLOW	F2	3.63	
FIRST EFFECT OVERHEAD FLOW	F5	1.81	
PRODUCT FLOW	F6	1.77	
SECOND EFFECT OVERHEAD FLOW	F7	1.68	
CONDENSER COOLING WATER FLOW	F9	40.13	
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	
PRODUCT CONCENTRATION	C6	0.090	
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.5	
STEAM TO FIRST EFFECT	T15	299.4	
SOLUTION IN FIRST EFFECT	T19	226.3	
VAPOR IN FIRST EFFECT	T2	224.2	
FIRST EFFECT STEAM CONDENSATE	T5	251.0	
SOLUTION TO SECOND EFFECT	T4	183.2	
STEAM TO SECOND EFFECT	T10	224.8	
PRODUCT	T34	158.4	
STEAM CONDENSATE SECOND EFFECT	T28	199.7	
SEPARATOR VAPOR	T12	160.2	
COOLING WATER INLET	T29	51.2	
COOLING WATER OUTLET	T1	94.6	
CONDENSER CONDENSATE	T11	143.7	
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		7.44	
SECOND EFFECT PRESSURE		-15.02	
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.78	
TOTAL COMPONENT BALANCE		1.37	

Table A-20 Steady State Data for Run DDC9 (-20% step in F)

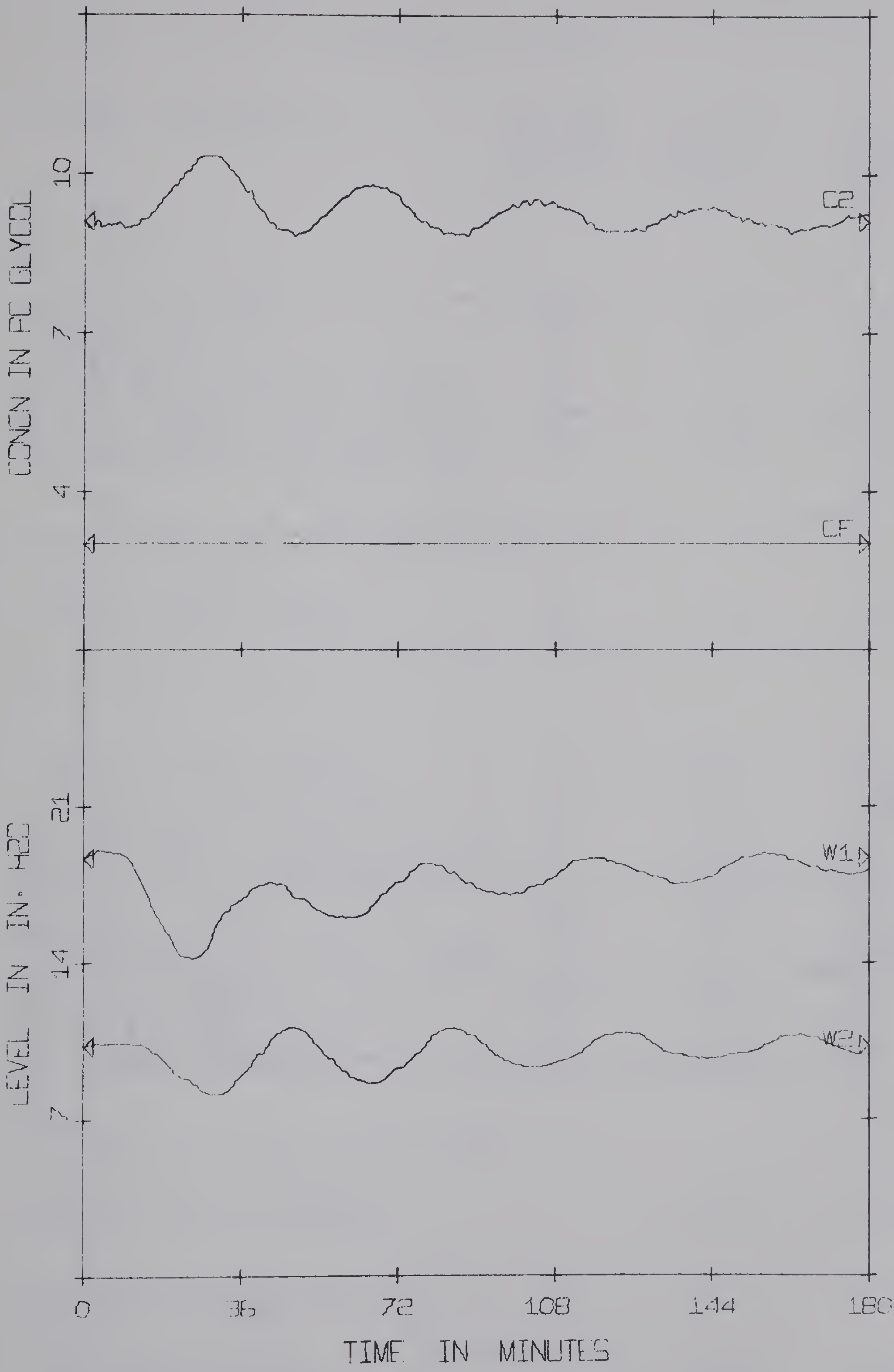


Figure A-19a Transient Data for Run DDC9 (-20% step in F)

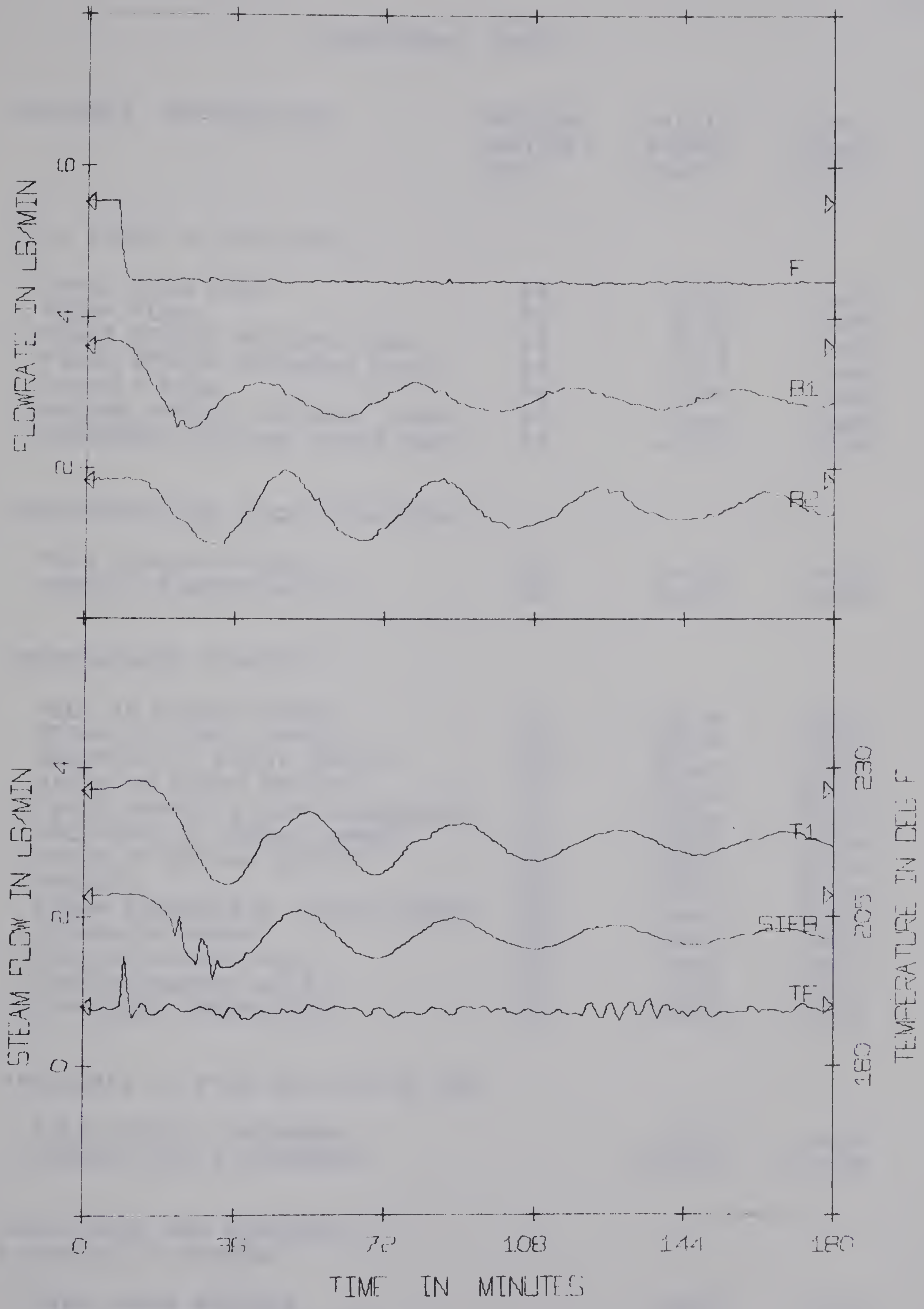


Figure A-19b Transient Data for Run DDC9 (-20% step in F)

EXPERIMENT DDC10

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	4.45
STEAM FLOW	F1	2.22	1.75
FIRST EFFECT BOTTOMS FLOW	F2	3.72	3.01
FIRST EFFECT OVERHEAD FLOW	F5	1.77	1.45
PRODUCT FLOW	F6	1.90	1.56
SECOND EFFECT OVERHEAD FLOW	F7	1.64	1.26
CONDENSER COOLING WATER FLOW	F9	40.50	39.89
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	187.8	189.3
STEAM TO FIRST EFFECT	T15	297.8	300.9
SOLUTION IN FIRST EFFECT	T19	224.7	221.5
VAPOR IN FIRST EFFECT	T2	223.2	220.1
FIRST EFFECT STEAM CONDENSATE	T5	250.6	242.0
SOLUTION TO SECOND EFFECT	T4	180.6	181.6
STEAM TO SECOND EFFECT	T10	223.7	220.6
PRODUCT	T34	157.3	158.4
STEAM CONDENSATE SECOND EFFECT	T28	196.0	190.0
SEPARATOR VAPOR	T12	159.1	161.4
COOLING WATER INLET	T29	53.2	49.1
COOLING WATER OUTLET	T1	97.5	87.1
CONDENSER CONDENSATE	T11	146.9	115.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		8.04	6.29
SECOND EFFECT PRESSURE		-14.06	-14.08
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.60	4.61
TOTAL COMPONENT BALANCE		-3.76	-4.77

Table A-21 Steady State Data for Run DDC10 (-20% step in F)

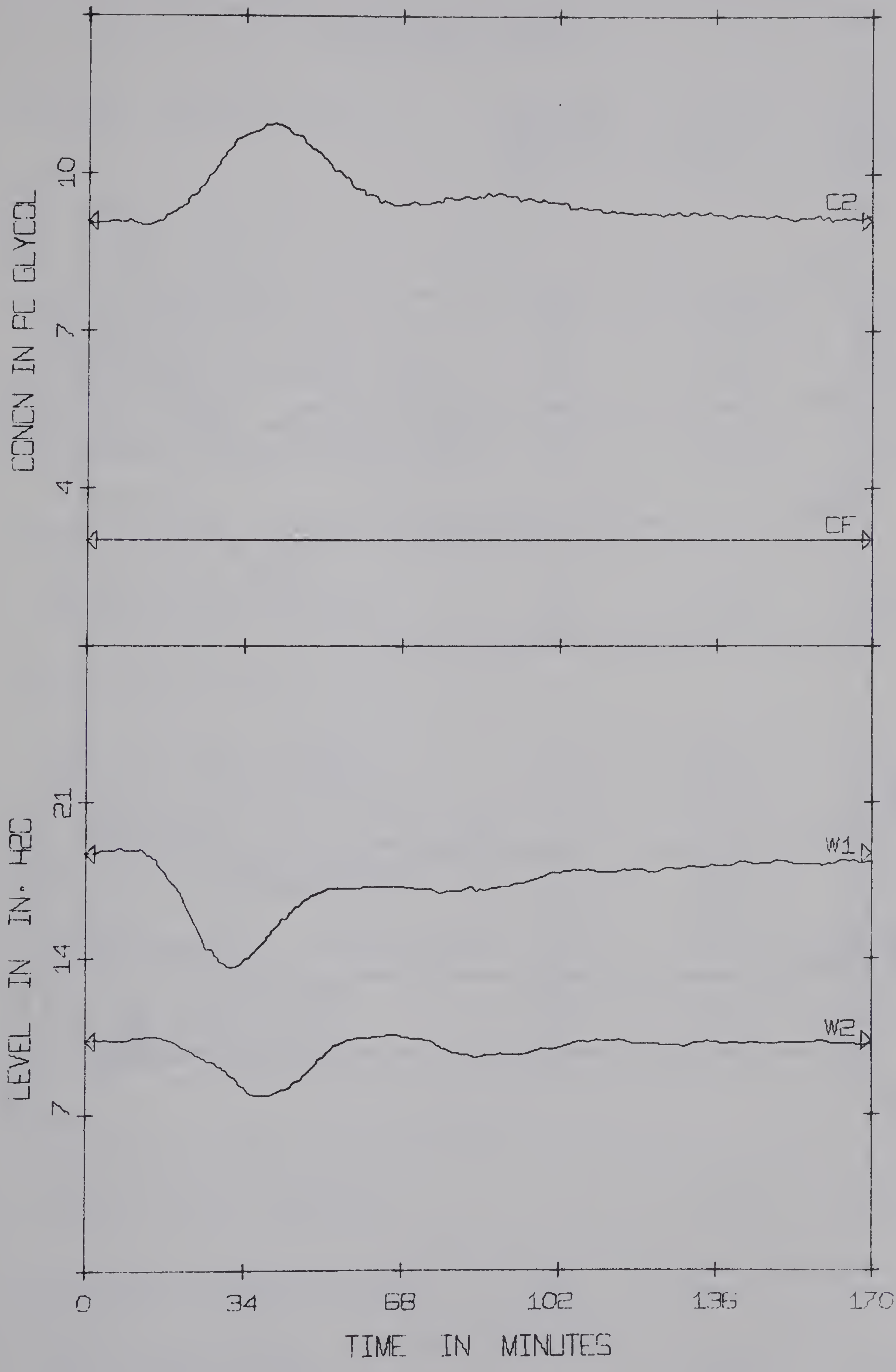


Figure A-20a Transient Data for Run DDC10 (-20% step in F)

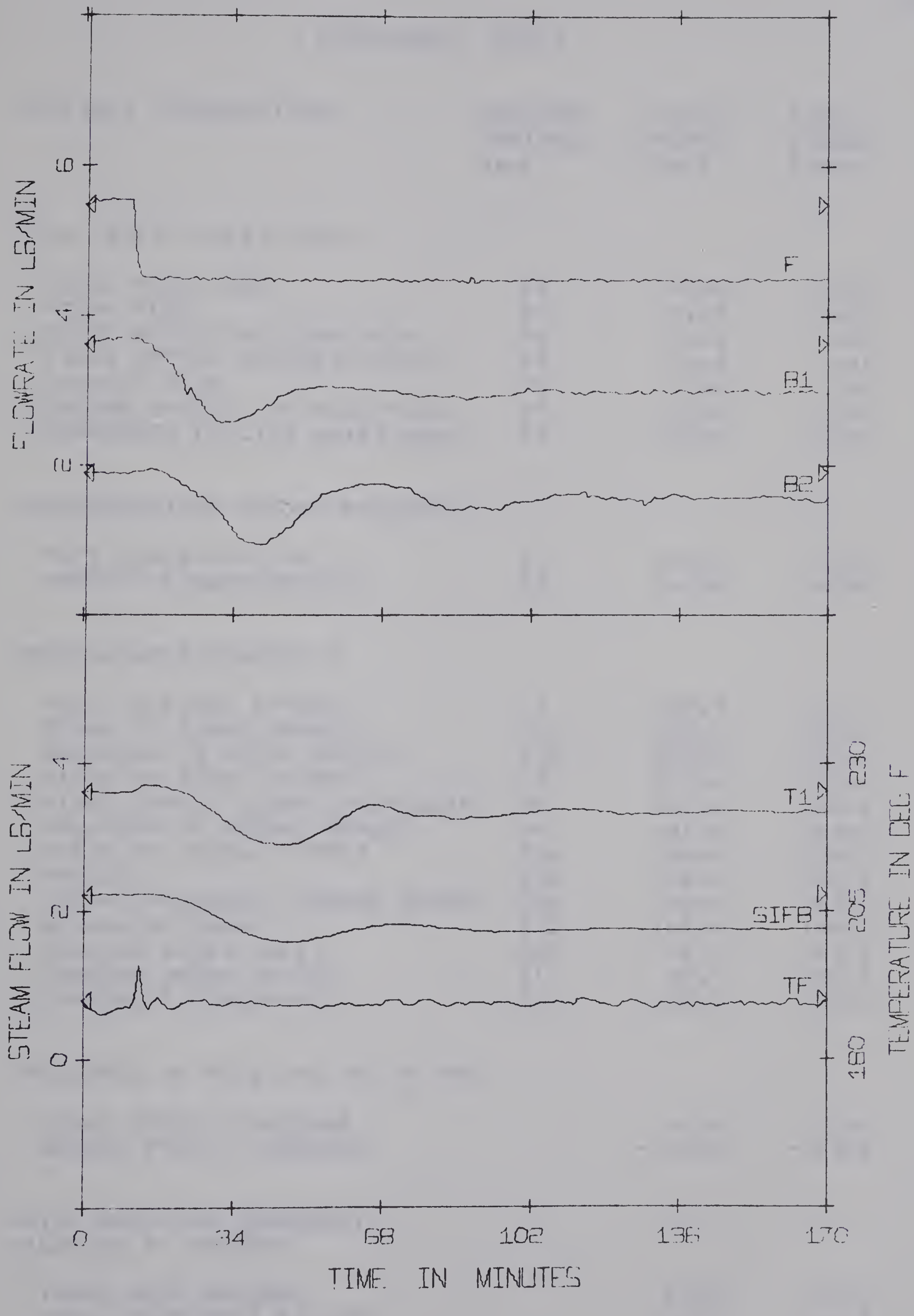


Figure A-20b Transient Data for Run DDC10 (-20% step in F)

EXPERIMENT DDC11

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.45	5.55
STEAM FLOW	F1	1.75	2.19
FIRST EFFECT BOTTOMS FLOW	F2	3.01	3.63
FIRST EFFECT OVERHEAD FLOW	F5	1.45	1.76
PRODUCT FLOW	F6	1.56	1.92
SECOND EFFECT OVERHEAD FLOW	F7	1.26	1.61
CONDENSER COOLING WATER FLOW	F9	39.89	40.26
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.3	188.7
STEAM TO FIRST EFFECT	T15	300.9	297.7
SOLUTION IN FIRST EFFECT	T19	221.5	224.3
VAPOR IN FIRST EFFECT	T2	220.1	222.4
FIRST EFFECT STEAM CONDENSATE	T5	242.0	248.9
SOLUTION TO SECOND EFFECT	T4	181.6	182.0
STEAM TO SECOND EFFECT	T10	220.6	223.1
PRODUCT	T34	158.4	158.6
STEAM CONDENSATE SECOND EFFECT	T28	190.0	197.5
SEPARATOR VAPOR	T12	161.4	160.1
COOLING WATER INLET	T29	49.1	48.2
COOLING WATER OUTLET	T1	87.1	92.9
CONDENSER CONDENSATE	T11	115.0	143.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.29	7.34
SECOND EFFECT PRESSURE		-14.08	-13.96
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.61	3.24
TOTAL COMPONENT BALANCE		-4.77	-4.23

Table A-22 Steady State Data for Run DDC11 (+20% step in F)

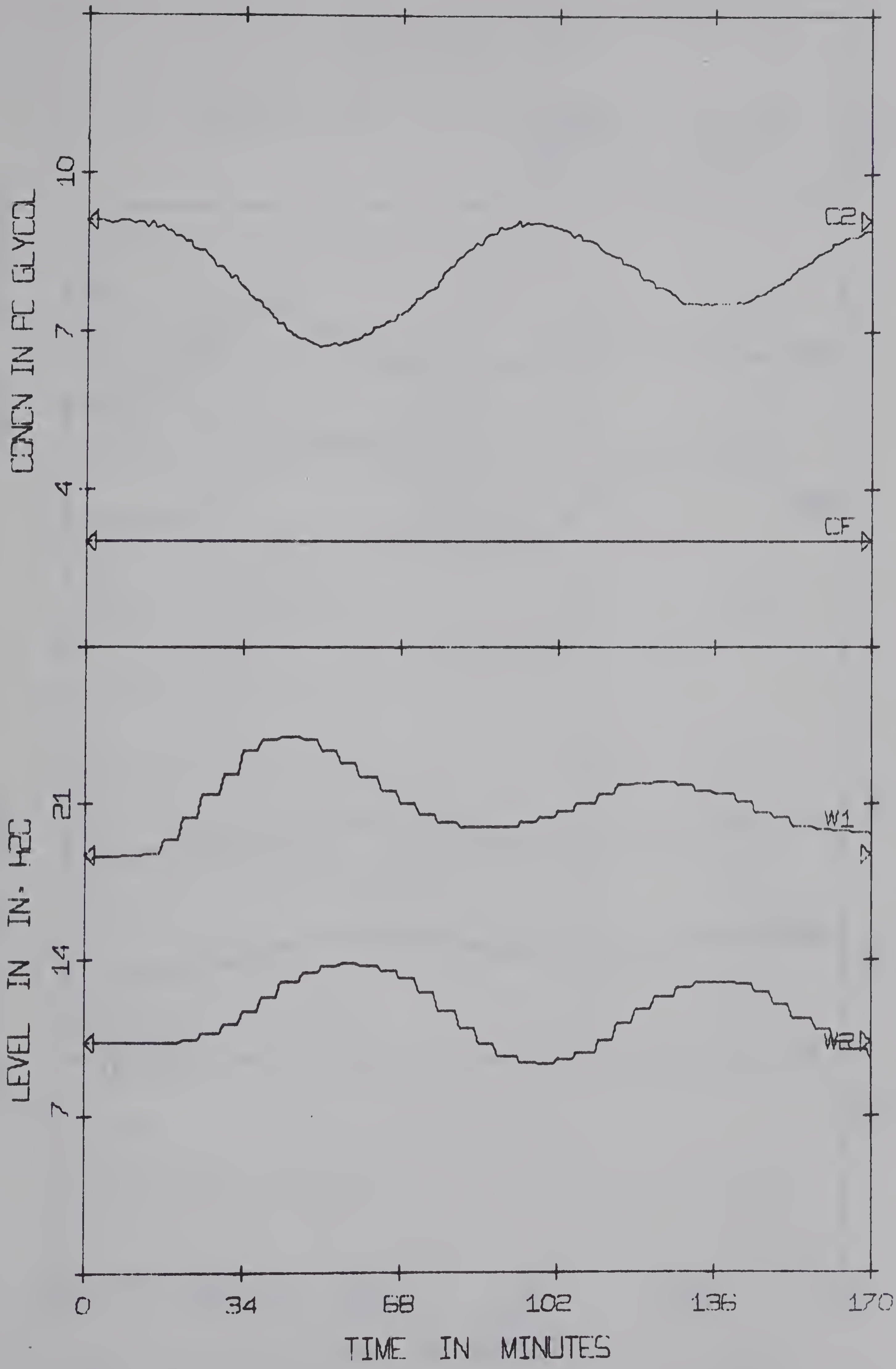


Figure A-21a Transient Data for Run DDC11 (+20% step in F)

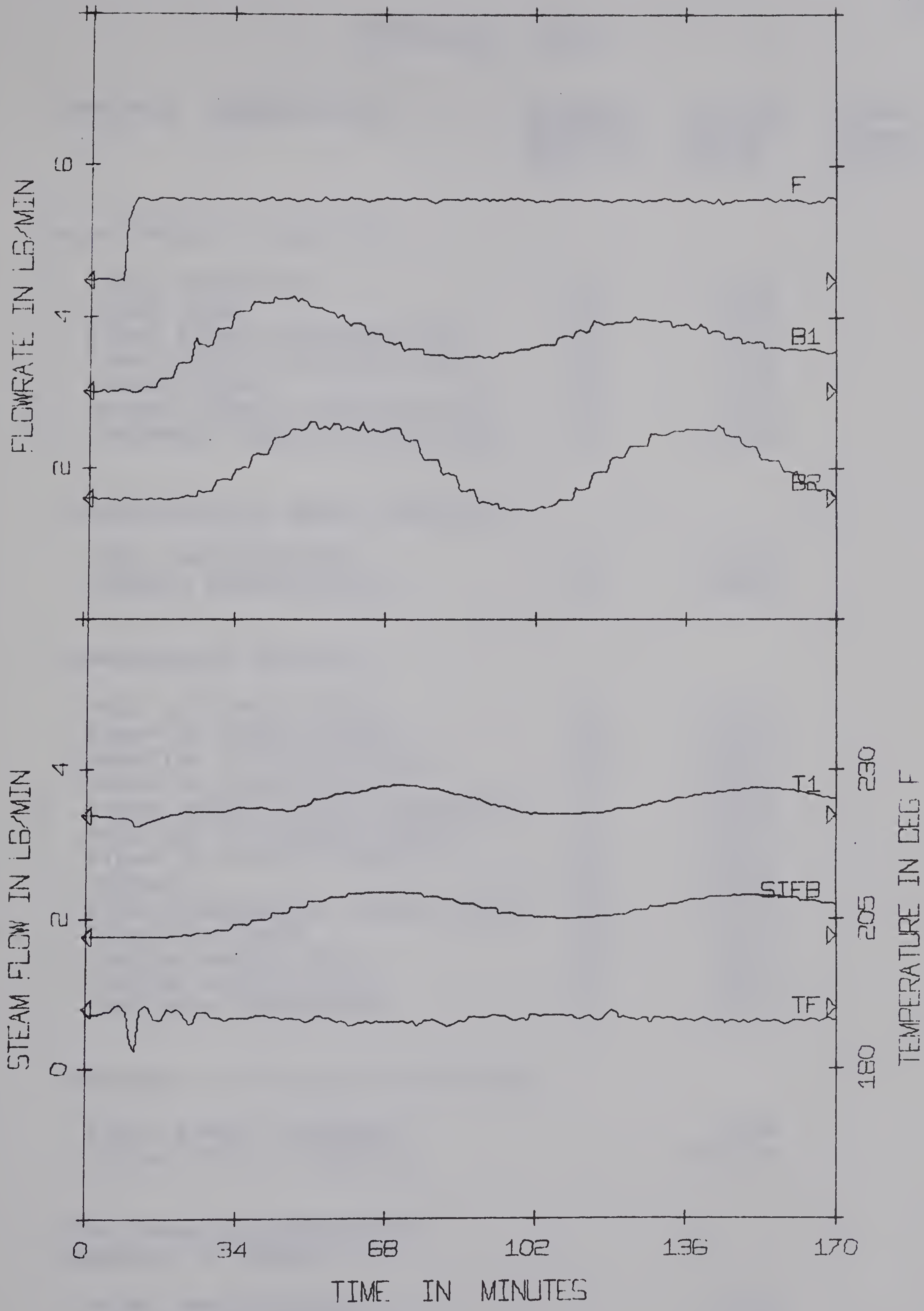


Figure A-21b Transient Data for Run DDC11 (+20% step in F)

EXPERIMENT DDC12

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.55	
STEAM FLOW	F1	2.19	
FIRST EFFECT BOTTOMS FLOW	F2	3.63	
FIRST EFFECT OVERHEAD FLOW	F5	1.76	
PRODUCT FLOW	F6	1.92	
SECOND EFFECT OVERHEAD FLOW	F7	1.61	
CONDENSER COOLING WATER FLOW	F9	40.26	
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	
PRODUCT CONCENTRATION	C6	0.090	
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.7	
STEAM TO FIRST EFFECT	T15	297.7	
SOLUTION IN FIRST EFFECT	T19	224.3	
VAPOR IN FIRST EFFECT	T2	222.4	
FIRST EFFECT STEAM CONDENSATE	T5	248.9	
SOLUTION TO SECOND EFFECT	T4	182.0	
STEAM TO SECOND EFFECT	T10	223.1	
PRODUCT	T34	158.6	
STEAM CONDENSATE SECOND EFFECT	T28	197.5	
SEPARATOR VAPOR	T12	160.1	
COOLING WATER INLET	T29	48.2	
COOLING WATER OUTLET	T1	92.9	
CONDENSER CONDENSATE	T11	143.1	
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		7.34	
SECOND EFFECT PRESSURE		-13.96	
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.24	
TOTAL COMPONENT BALANCE		-4.23	

Table A-23 Steady State Data for Run DDC12 (-20% step in F)

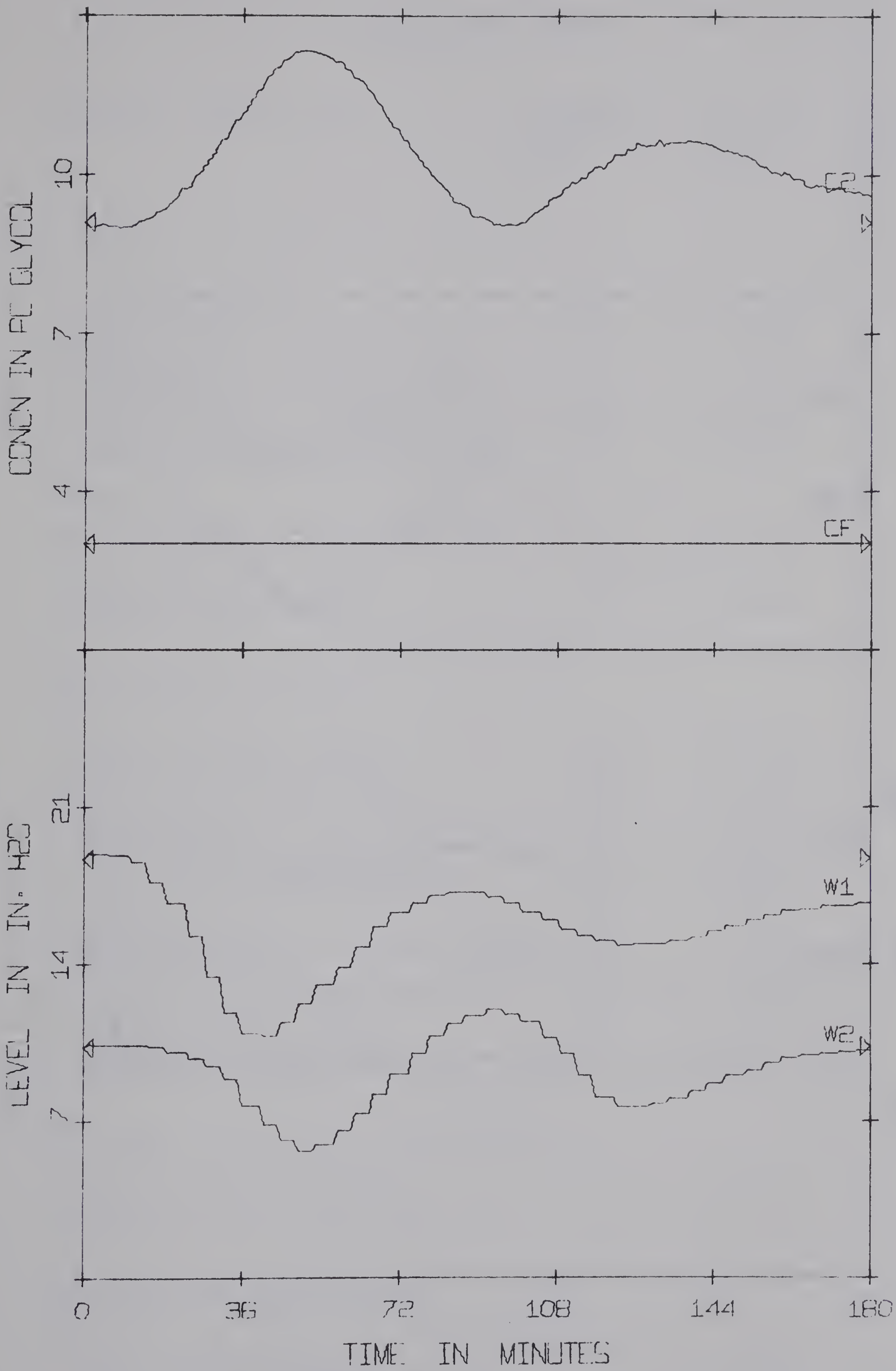


Figure A-22a Transient Data for Run DDC12 (-20% step in F)

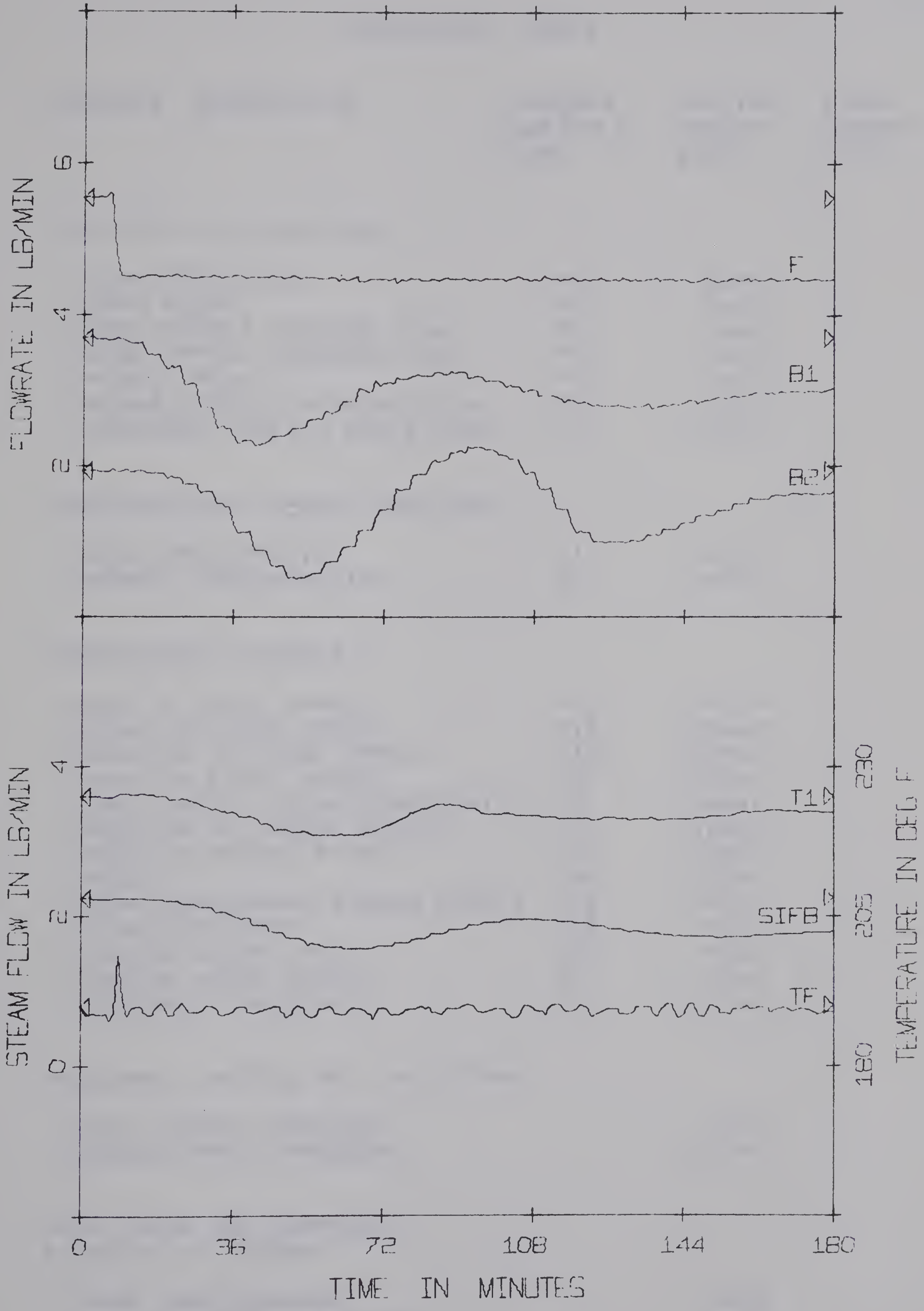
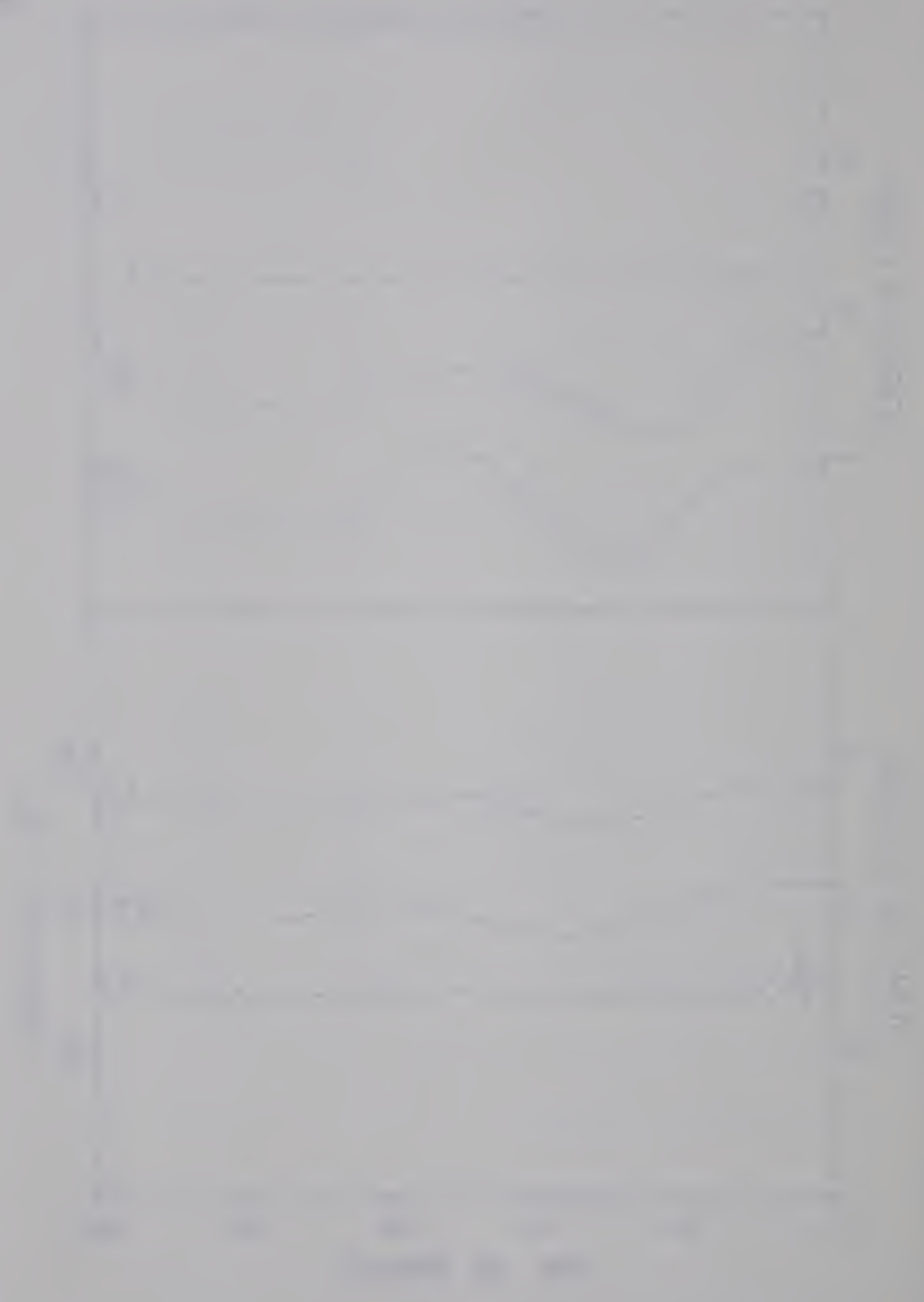


Figure A-22b Transient Data for Run DDC12 (-20% step in F)



EXPERIMENT DDC13

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	
STEAM FLOW	F1	2.28	
FIRST EFFECT BOTTOMS FLOW	F2	3.60	
FIRST EFFECT OVERHEAD FLOW	F5	1.83	
PRODUCT FLOW	F6	1.77	
SECOND EFFECT OVERHEAD FLOW	F7	1.69	
CONDENSER COOLING WATER FLOW	F9	39.95	
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	
PRODUCT CONCENTRATION	C6	0.090	
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	187.9	
STEAM TO FIRST EFFECT	T15	296.7	
SOLUTION IN FIRST EFFECT	T19	224.9	
VAPOR IN FIRST EFFECT	T2	222.8	
FIRST EFFECT STEAM CONDENSATE	T5	248.4	
SOLUTION TO SECOND EFFECT	T4	182.1	
STEAM TO SECOND EFFECT	T10	223.1	
PRODUCT	T34	157.0	
STEAM CONDENSATE SECOND EFFECT	T28	197.4	
SEPARATOR VAPOR	T12	158.9	
COOLING WATER INLET	T29	47.9	
COOLING WATER OUTLET	T1	92.2	
CONDENSER CONDENSATE	T11	141.9	
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		7.97	
SECOND EFFECT PRESSURE		-13.52	
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.43	
TOTAL COMPONENT BALANCE		4.32	

Table A-24 Steady State Data for Run DDC13 (-20% step in F)

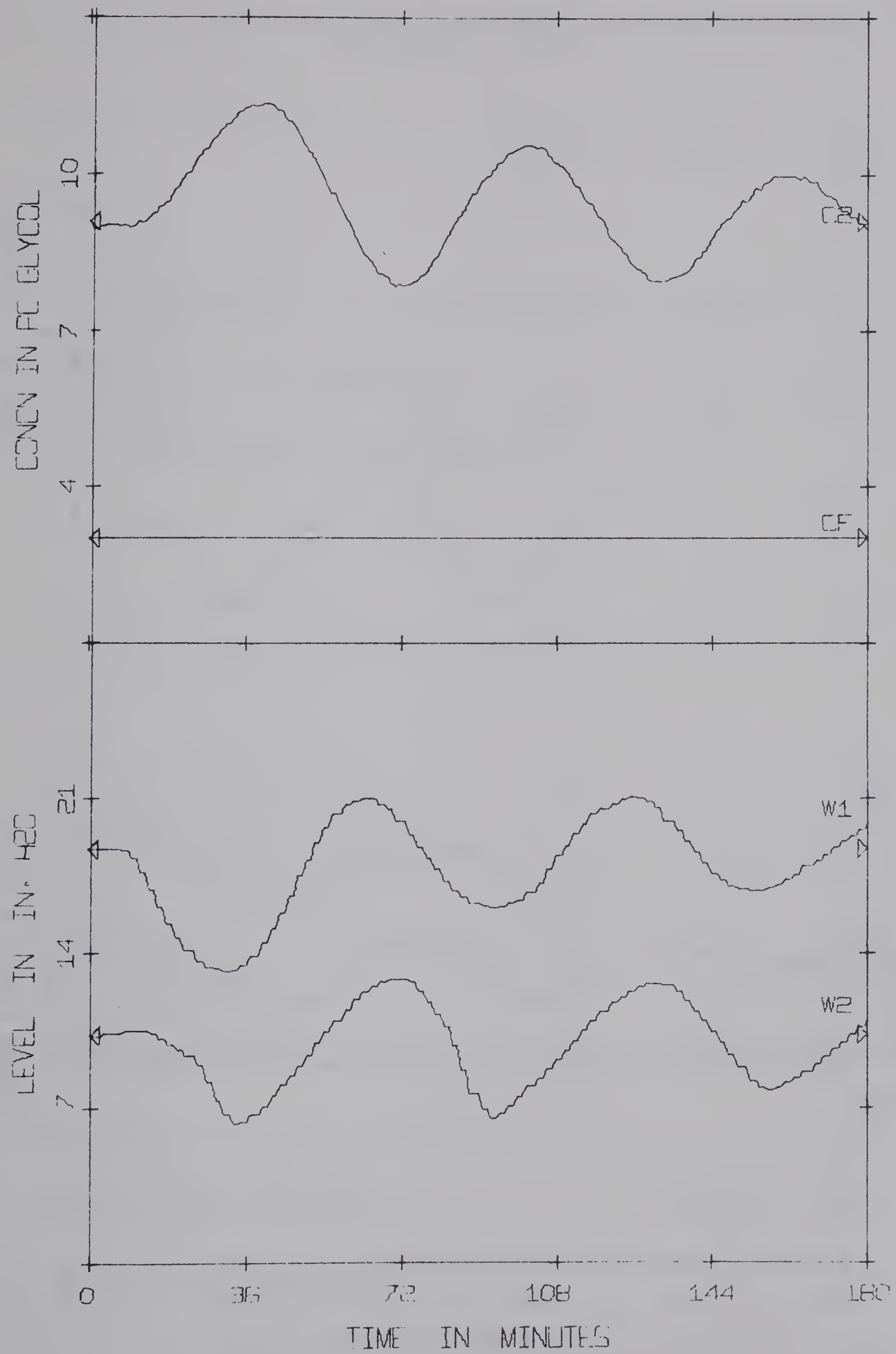


Figure A-23a Transient Data for Run DDC13 (-20% step in F)

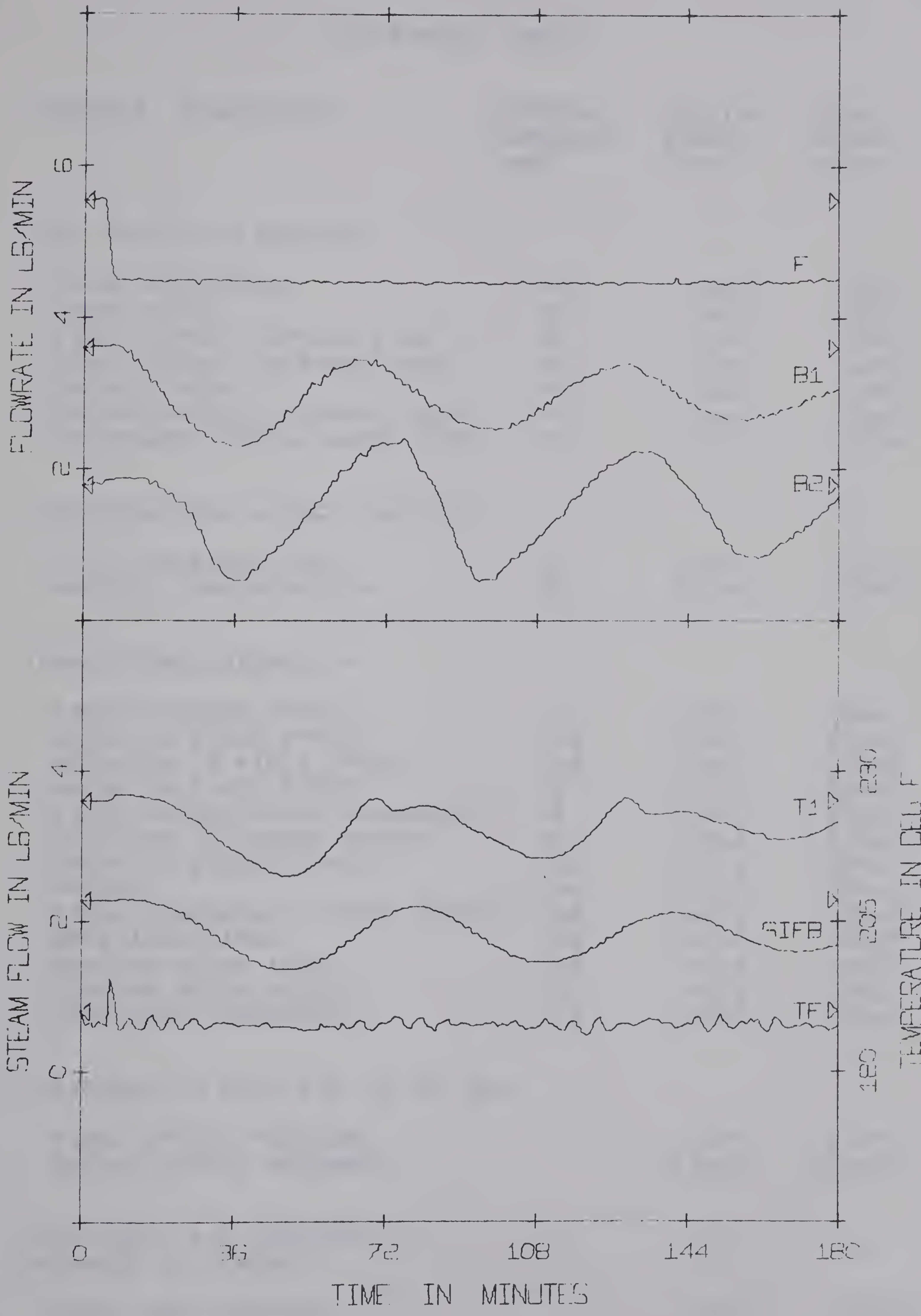


Figure A-23b Transient Data for Run DDC13 (-20% step in F)

EXPERIMENT DDC14

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.50	5.54
STEAM FLOW	F1	1.77	2.32
FIRST EFFECT BOTTOMS FLOW	F2	2.91	3.61
FIRST EFFECT OVERHEAD FLOW	F5	1.47	1.85
PRODUCT FLOW	F6	1.42	1.85
SECOND EFFECT OVERHEAD FLOW	F7	1.26	1.63
CONDENSER COOLING WATER FLOW	F9	41.01	41.06
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	187.2	186.7
STEAM TO FIRST EFFECT	T15	298.2	294.1
SOLUTION IN FIRST EFFECT	T19	218.3	225.9
VAPOR IN FIRST EFFECT	T2	217.0	224.2
FIRST EFFECT STEAM CONDENSATE	T5	239.1	251.1
SOLUTION TO SECOND EFFECT	T4	178.2	181.4
STEAM TO SECOND EFFECT	T10	217.3	224.7
PRODUCT	T34	154.7	159.2
STEAM CONDENSATE SECOND EFFECT	T28	188.2	199.8
SEPARATOR VAPOR	T12	157.1	160.9
COOLING WATER INLET	T29	49.6	49.3
COOLING WATER OUTLET	T1	87.3	96.1
CONDENSER CONDENSATE	T11	118.0	144.9
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		5.84	10.76
SECOND EFFECT PRESSURE		-13.37	-13.40
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.68	2.69
TOTAL COMPONENT BALANCE		1.66	-0.05

Table A-25 Steady State Data for Run DDC14 (+20% step in F)

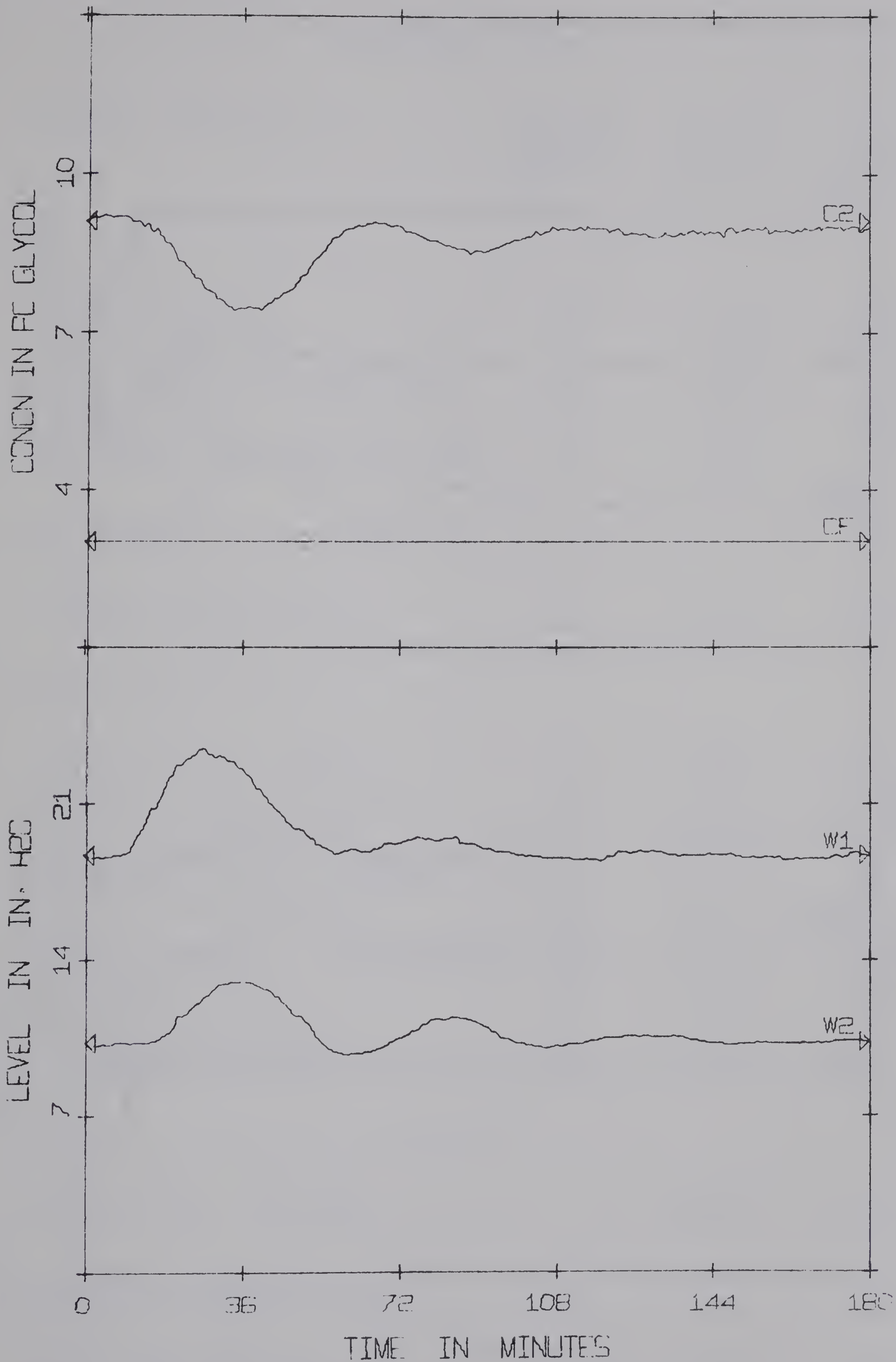


Figure A-24a Transient Data for Run DDC14 (+20% step in F)

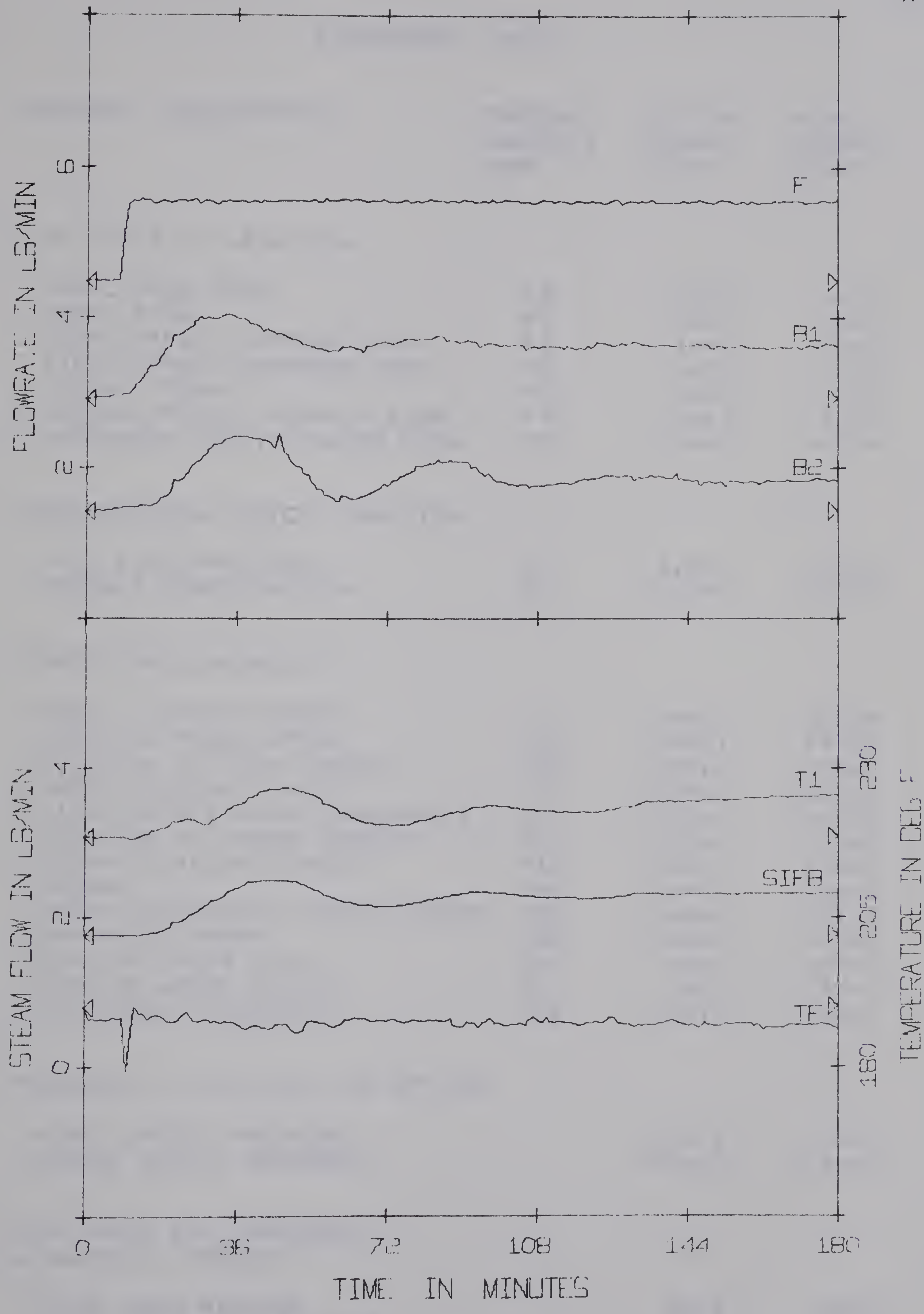


Figure A-24b Transient Data for Run DDC14 (+20% step in F)

EXPERIMENT DDC15

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	4.46
STEAM FLOW	F1	2.32	1.87
FIRST EFFECT BOTTOMS FLOW	F2	3.61	2.92
FIRST EFFECT OVERHEAD FLOW	F5	1.85	1.53
PRODUCT FLOW	F6	1.85	1.87
SECOND EFFECT OVERHEAD FLOW	F7	1.63	1.27
CONDENSER COOLING WATER FLOW	F9	41.06	40.29
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	186.7	187.8
STEAM TO FIRST EFFECT	T15	294.1	297.8
SOLUTION IN FIRST EFFECT	T19	225.9	221.9
VAPOR IN FIRST EFFECT	T2	224.2	220.2
FIRST EFFECT STEAM CONDENSATE	T5	251.1	242.6
SOLUTION TO SECOND EFFECT	T4	181.4	185.2
STEAM TO SECOND EFFECT	T10	224.7	220.7
PRODUCT	T34	159.2	164.2
STEAM CONDENSATE SECOND EFFECT	T28	199.8	197.3
SEPARATOR VAPOR	T12	160.9	167.8
COOLING WATER INLET	T29	49.3	49.5
COOLING WATER OUTLET	T1	96.1	88.3
CONDENSER CONDENSATE	T11	144.9	124.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		10.76	7.48
SECOND EFFECT PRESSURE		-13.40	-13.46
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.69	3.98
TOTAL COMPONENT BALANCE		-0.05	1.47

Table A-26 Steady State Data for Run DDC15 (-20% step in F)

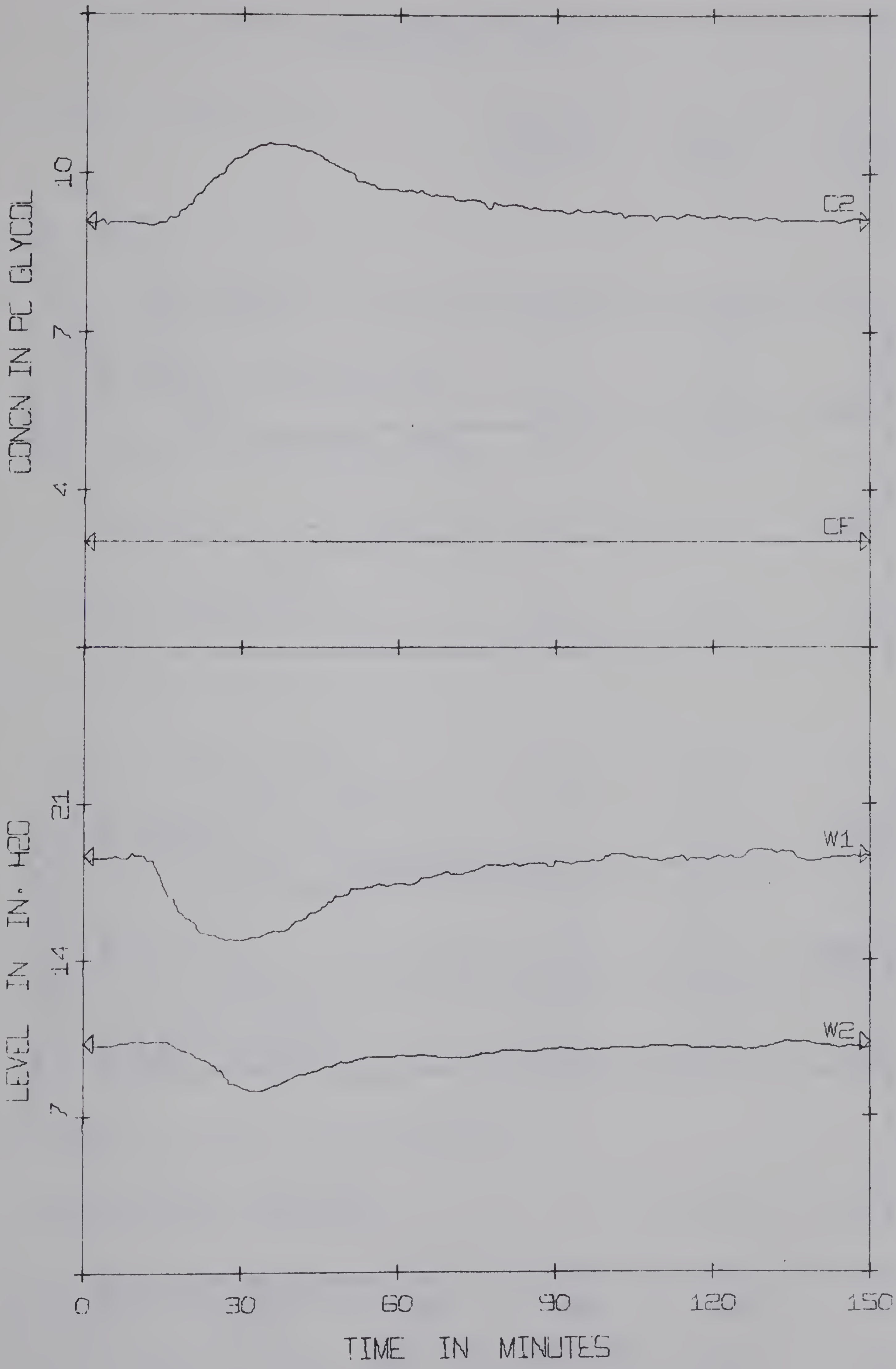


Figure A-25a Transient Data for Run DDC15 (-20% step in F)

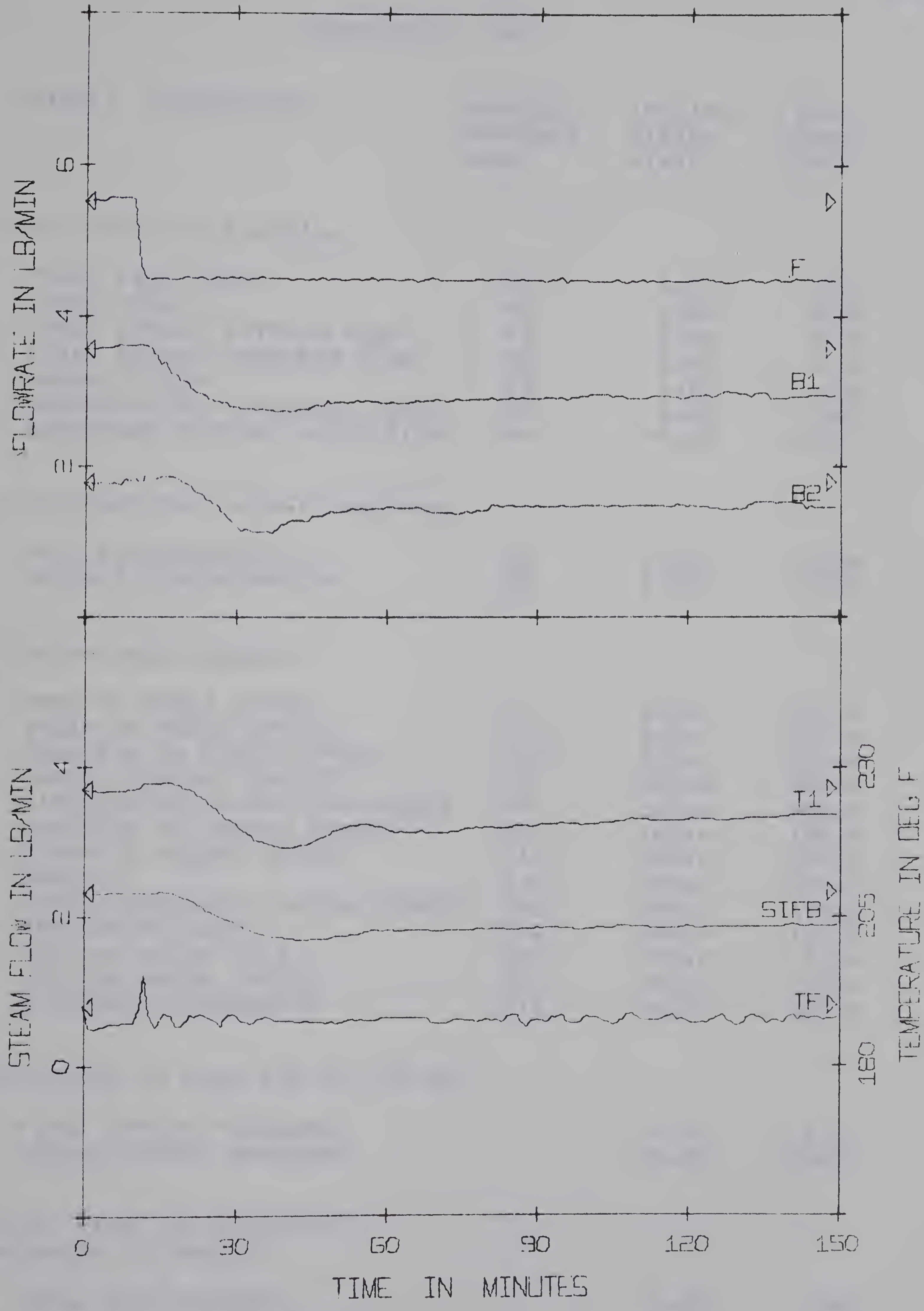
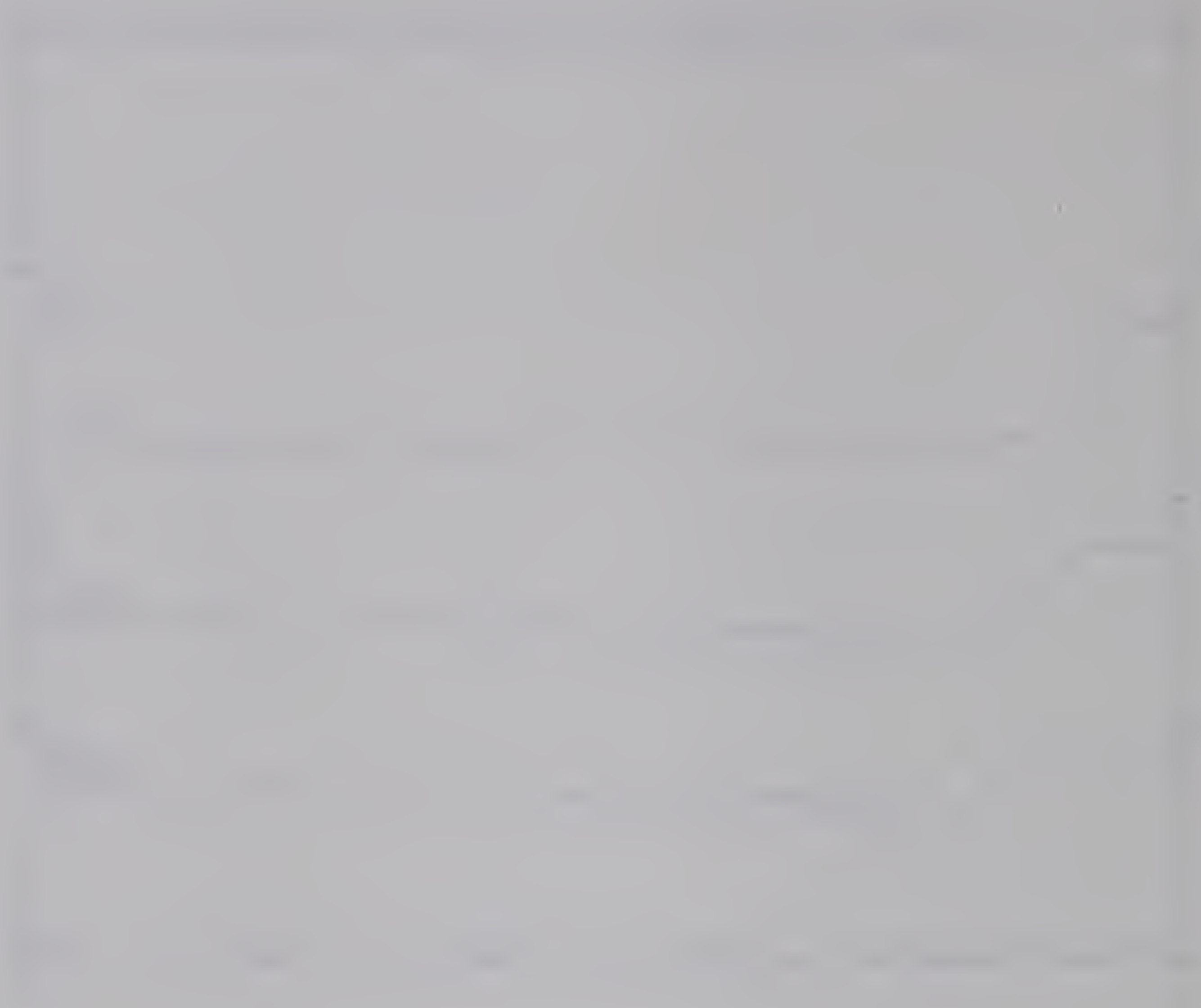


Figure A-25b Transient Data for Run DDC15 (-20% step in F)



EXPERIMENT INF1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.51	5.55
STEAM FLOW	F1	1.80	2.29
FIRST EFFECT BOTTOMS FLOW	F2	2.96	3.62
FIRST EFFECT OVERHEAD FLOW	F5	1.47	1.82
PRODUCT FLOW	F6	1.50	1.86
SECOND EFFECT OVERHEAD FLOW	F7	1.22	1.61
CONDENSER COOLING WATER FLOW	F9	39.83	39.93
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	187.5	188.0
STEAM TO FIRST EFFECT	T15	301.1	297.5
SOLUTION IN FIRST EFFECT	T19	222.5	225.5
VAPOR IN FIRST EFFECT	T2	221.0	224.0
FIRST EFFECT STEAM CONDENSATE	T5	242.8	250.3
SOLUTION TO SECOND EFFECT	T4	181.4	183.4
STEAM TO SECOND EFFECT	T10	221.6	224.5
PRODUCT	T34	158.1	158.6
STEAM CONDENSATE SECOND EFFECT	T28	195.0	199.6
SEPARATOR VAPOR	T12	160.8	160.4
COOLING WATER INLET	T29	52.8	52.6
COOLING WATER OUTLET	T1	90.5	97.6
CONDENSER CONDENSATE	T11	141.2	148.3
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.40	8.15
SECOND EFFECT PRESSURE		-15.85	-15.82
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.65	3.92
TOTAL COMPONENT BALANCE		-1.53	2.84

Table A-27 Steady State Data for Run INF1 (+20% step in F)

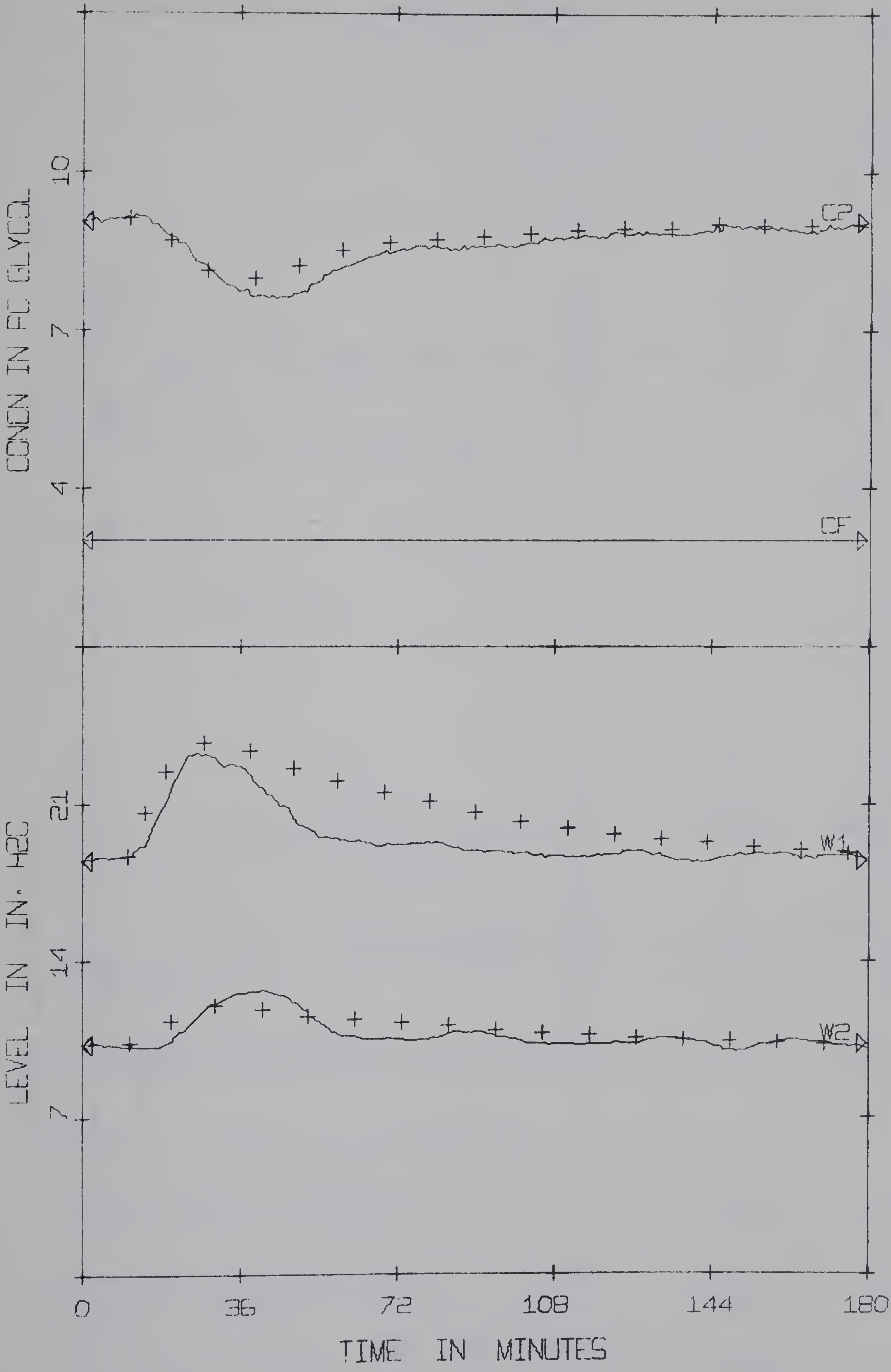


Figure A-26a Transient Data for Run INF1 (+20% step in F)

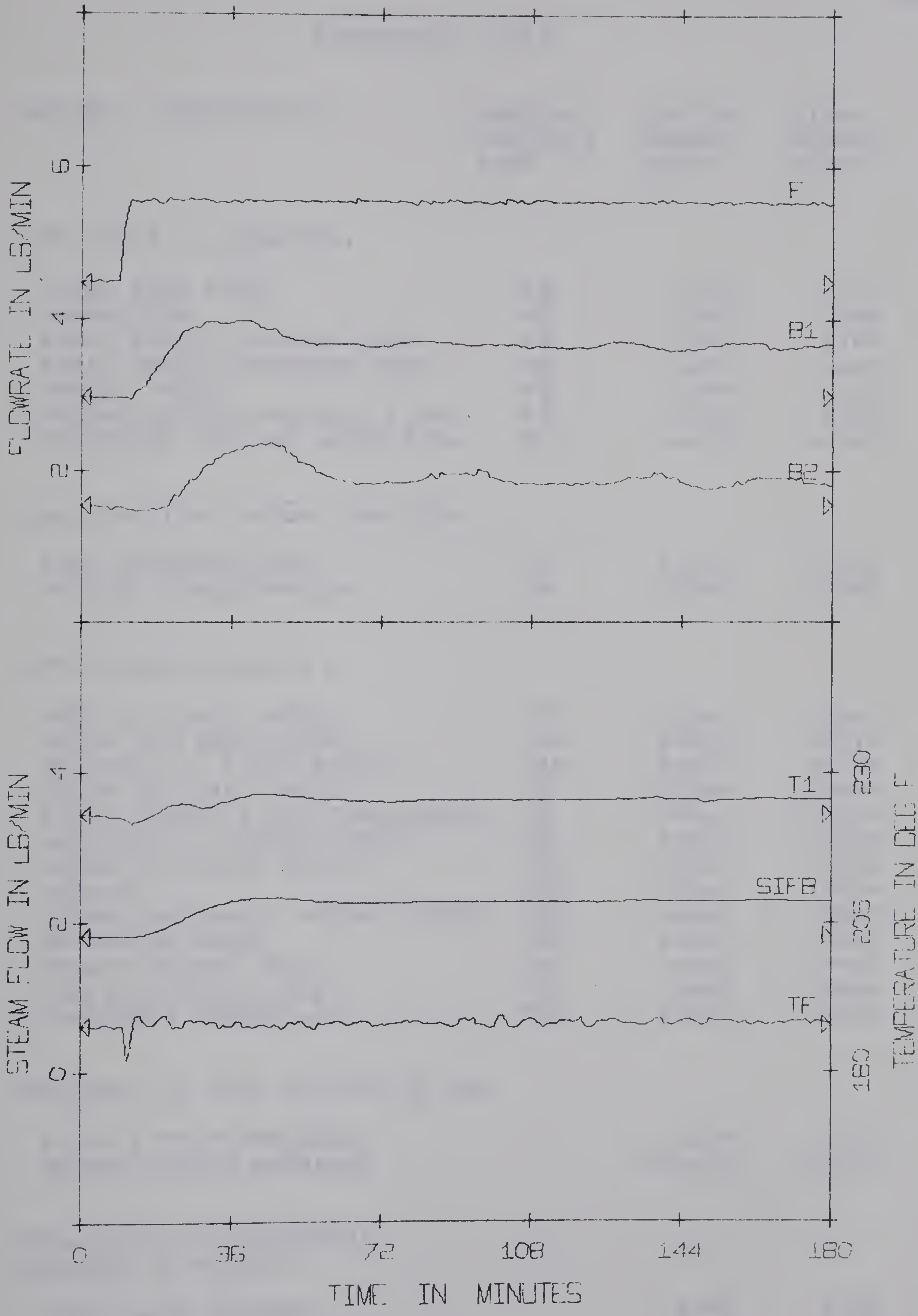


Figure A-26b Transient Data for Run INF1 (+20% step in F)

EXPERIMENT INF2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.50	4.47
STEAM FLOW	F1	1.81	1.96
FIRST EFFECT BOTTOMS FLOW	F2	2.98	2.86
FIRST EFFECT OVERHEAD FLOW	F5	1.47	1.61
PRODUCT FLOW	F6	1.49	1.21
SECOND EFFECT OVERHEAD FLOW	F7	1.30	1.52
CONDENSER COOLING WATER FLOW	F9	40.13	40.01
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.025
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.3	189.1
STEAM TO FIRST EFFECT	T15	299.0	297.8
SOLUTION IN FIRST EFFECT	T19	221.9	221.3
VAPOR IN FIRST EFFECT	T2	220.4	214.8
FIRST EFFECT STEAM CONDENSATE	T5	242.9	243.9
SOLUTION TO SECOND EFFECT	T4	177.7	177.0
STEAM TO SECOND EFFECT	T10	221.2	220.2
PRODUCT	T34	151.6	150.4
STEAM CONDENSATE SECOND EFFECT	T28	186.7	188.7
SEPARATOR VAPOR	T12	154.1	153.3
COOLING WATER INLET	T29	48.8	48.7
COOLING WATER OUTLET	T1	86.8	88.6
CONDENSER CONDENSATE	T11	113.6	124.8
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.19	6.17
SECOND EFFECT PRESSURE		-15.85	-12.25
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.65	4.03
TOTAL COMPONENT BALANCE		-1.53	3.92

Table A-28 Steady State Data for Run INF2 (-16.7% step in CF)

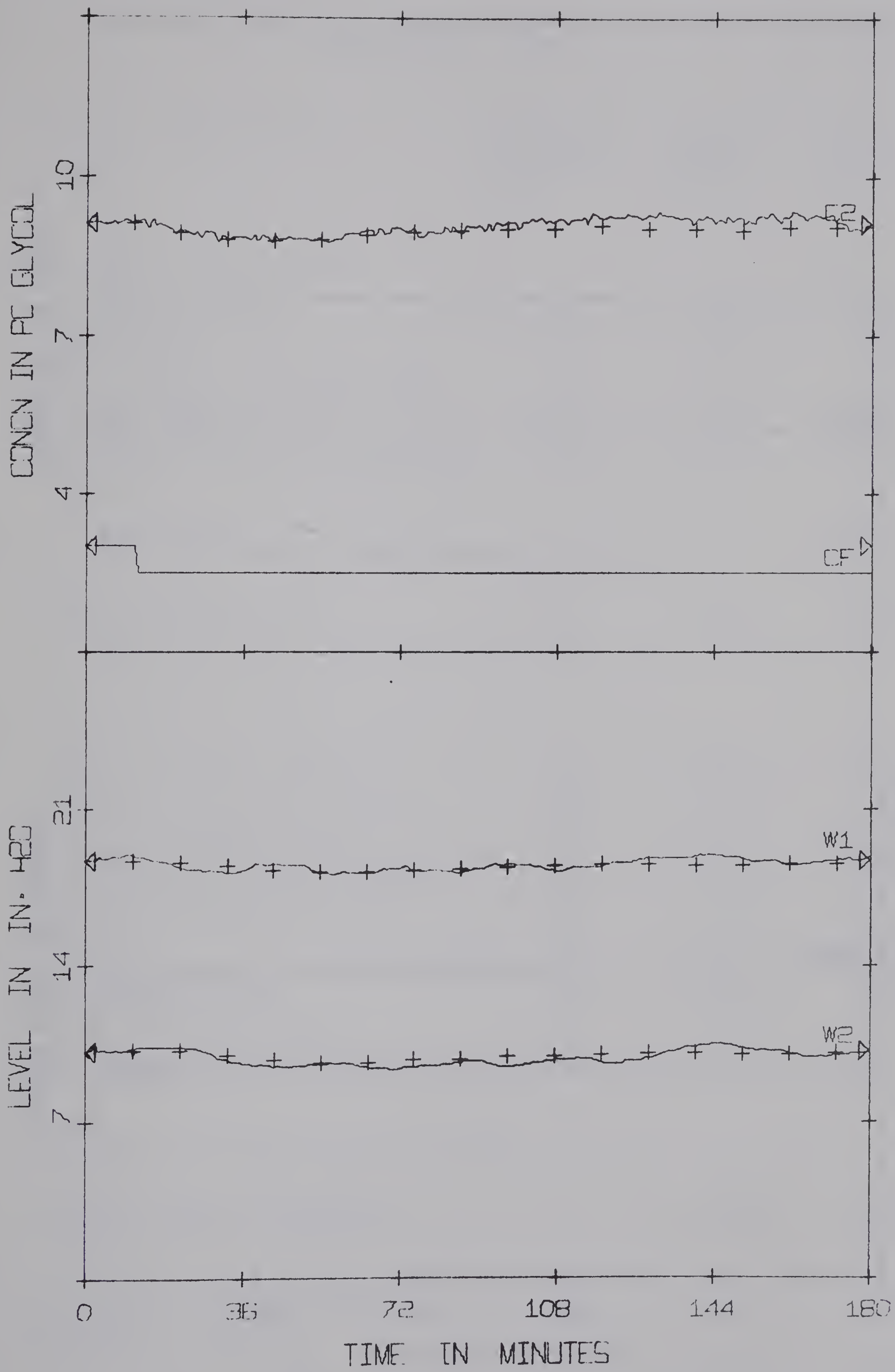
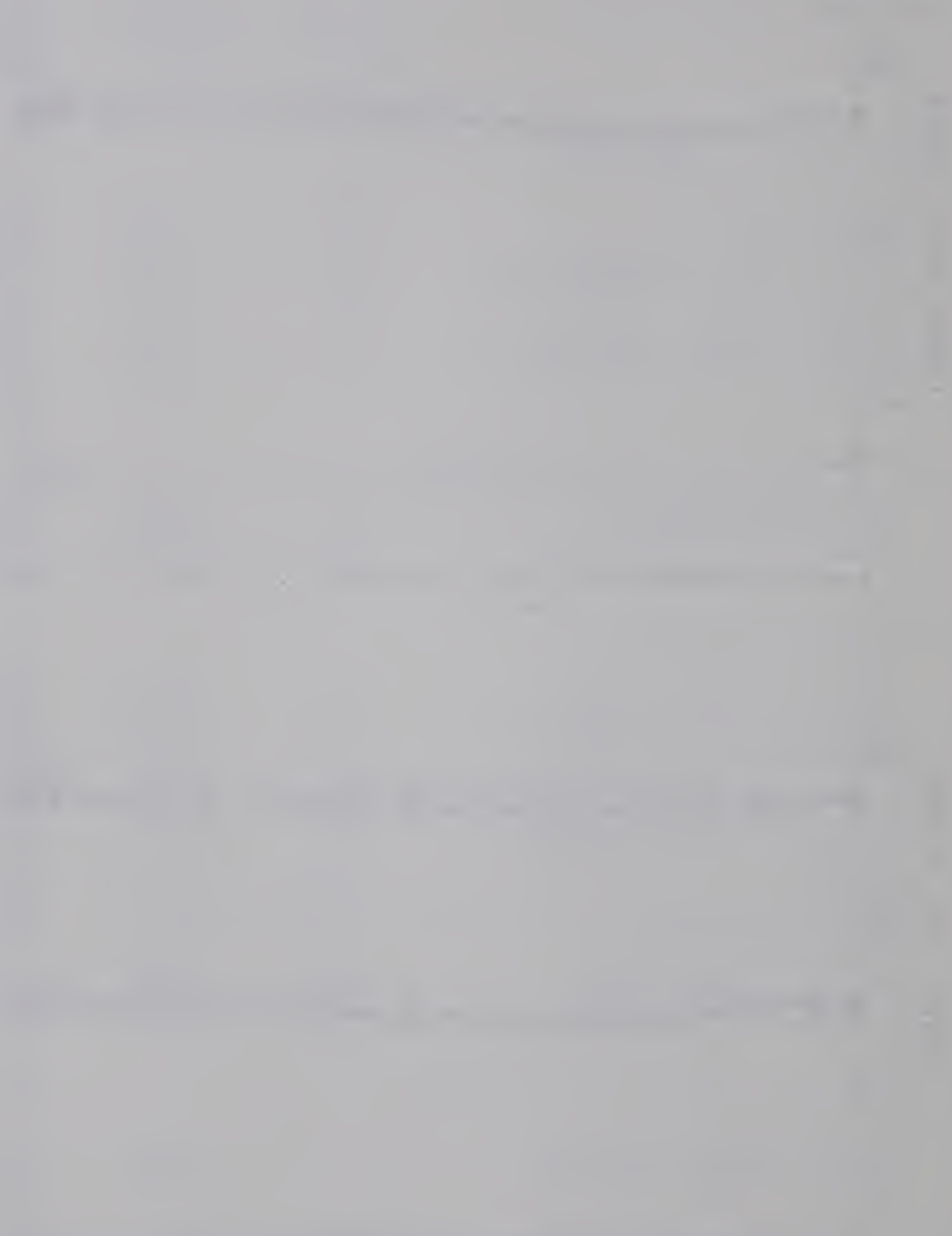


Figure A-27a Transient Data for Run INF2 (-16.7% step in CF)



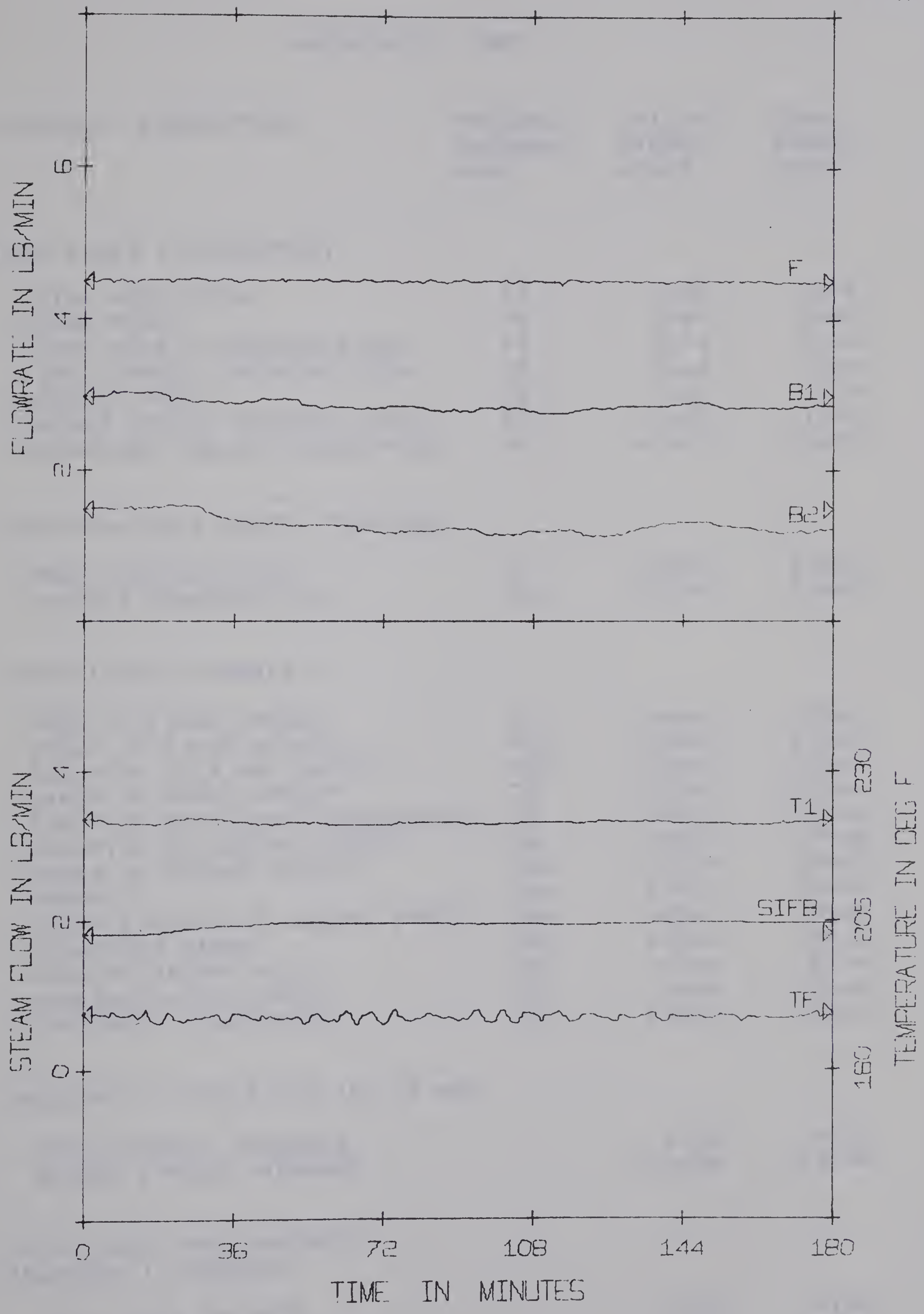


Figure A-27b Transient Data for Run INF2 (-16.7% step in CF)

EXPERIMENT INF3

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	4.48
STEAM FLOW	F1	1.83	2.00
FIRST EFFECT BOTTOMS FLOW	F2	2.95	2.86
FIRST EFFECT OVERHEAD FLOW	F5	1.48	1.64
PRODUCT FLOW	F6	1.44	1.19
SECOND EFFECT OVERHEAD FLOW	F7	1.20	1.27
CONDENSER COOLING WATER FLOW	F9	41.87	41.78
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.025
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.2	190.2
STEAM TO FIRST EFFECT	T15	298.8	297.6
SOLUTION IN FIRST EFFECT	T19	222.0	221.7
VAPOR IN FIRST EFFECT	T2	220.3	219.8
FIRST EFFECT STEAM CONDENSATE	T5	242.4	243.2
SOLUTION TO SECOND EFFECT	T4	180.0	179.8
STEAM TO SECOND EFFECT	T10	220.8	220.3
PRODUCT	T34	155.7	154.9
STEAM CONDENSATE SECOND EFFECT	T28	189.7	193.3
SEPARATOR VAPOR	T12	158.5	158.2
COOLING WATER INLET	T29	47.6	52.4
COOLING WATER OUTLET	T1	84.8	91.0
CONDENSER CONDENSATE	T11	100.0	118.9
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.52	6.32
SECOND EFFECT PRESSURE		-13.98	-13.98
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.72	5.40
TOTAL COMPONENT BALANCE		0.62	3.68

Table A-29 Steady State Data for Run INF3 (-16.7% step in CF)

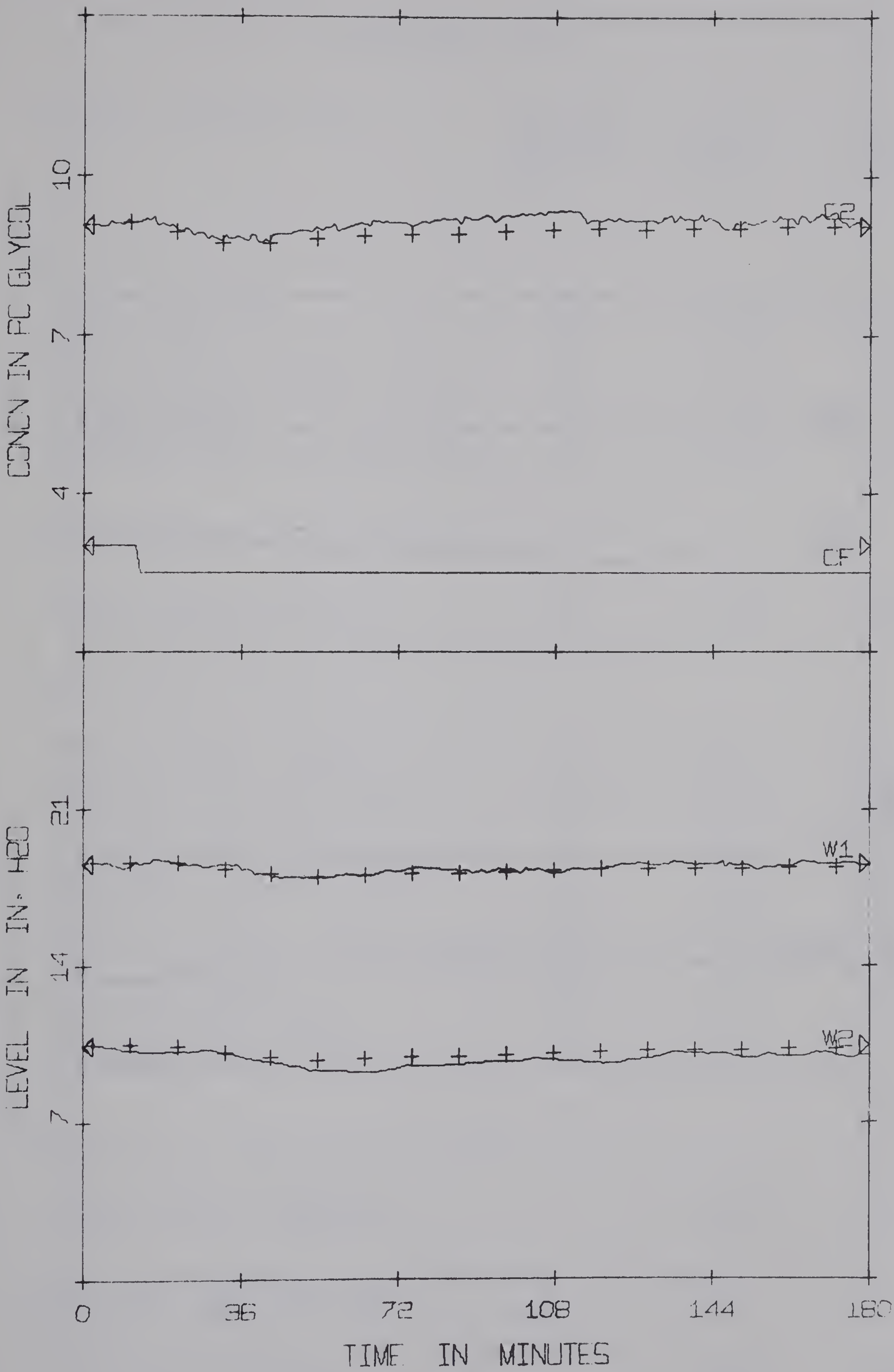


Figure A-28a Transient Data for Run INF3 (-16.7% step in CF)

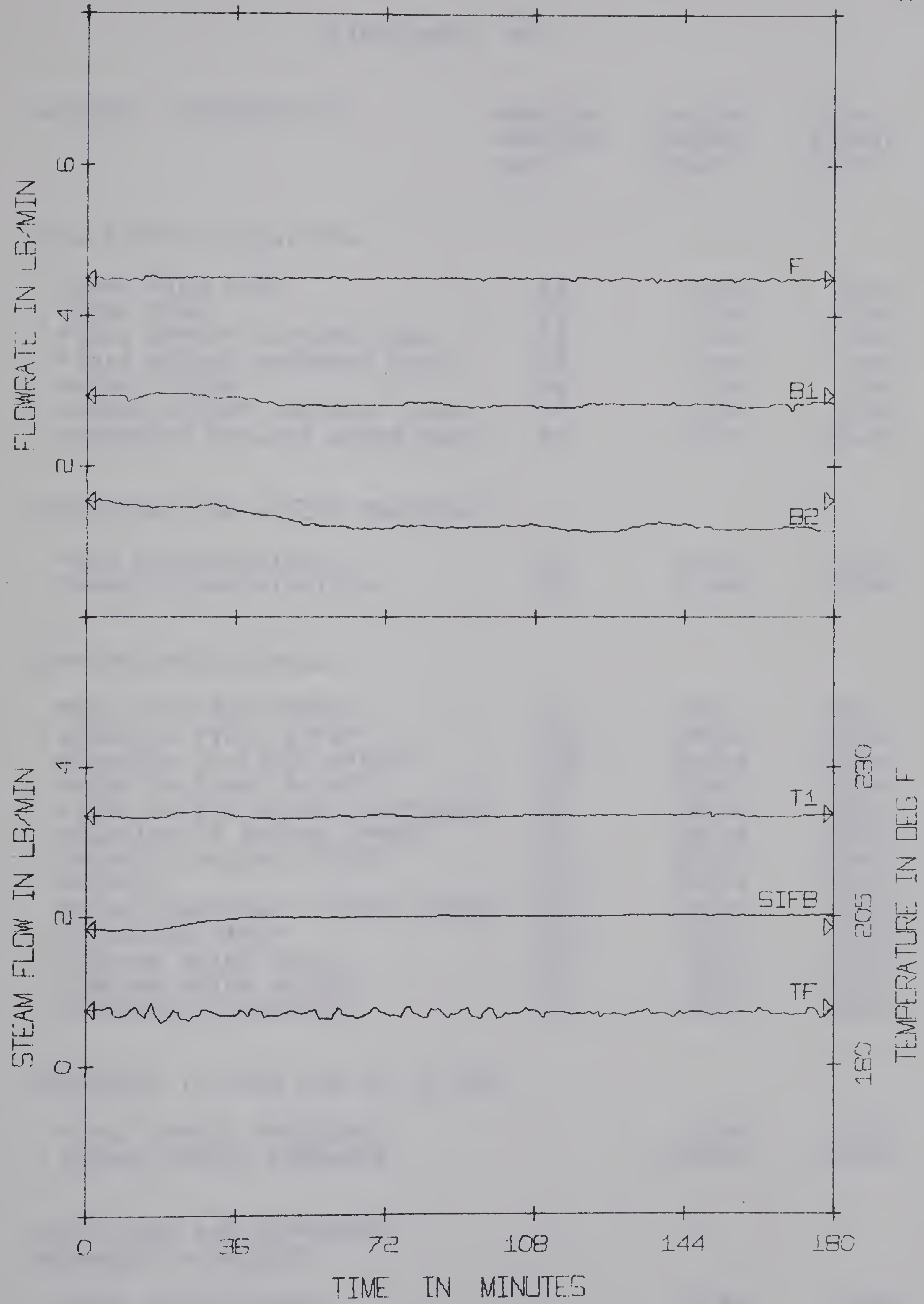


Figure A-28b Transient Data for Run INF3 (-16.7% step in CF)

EXPERIMENT FF1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	5.54
STEAM FLOW	F1	1.80	2.30
FIRST EFFECT BOTTOMS FLOW	F2	2.95	3.75
FIRST EFFECT OVERHEAD FLOW	F5	1.44	1.81
PRODUCT FLOW	F6	1.50	1.85
SECOND EFFECT OVERHEAD FLOW	F7	1.29	1.70
CONDENSER COOLING WATER FLOW	F9	39.91	39.95
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.089	0.089
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.1	189.7
STEAM TO FIRST EFFECT	T15	299.2	297.0
SOLUTION IN FIRST EFFECT	T19	216.8	227.4
VAPOR IN FIRST EFFECT	T2	215.1	225.8
FIRST EFFECT STEAM CONDENSATE	T5	238.0	253.7
SOLUTION TO SECOND EFFECT	T4	181.8	182.5
STEAM TO SECOND EFFECT	T10	215.5	226.1
PRODUCT	T34	157.8	158.3
STEAM CONDENSATE SECOND EFFECT	T28	191.5	198.2
SEPARATOR VAPOR	T12	161.4	160.5
COOLING WATER INLET	T29	53.6	52.7
COOLING WATER OUTLET	T1	90.5	97.7
CONDENSER CONDENSATE	T11	141.0	146.8
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.48	9.38
SECOND EFFECT PRESSURE		-15.51	-15.48
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.84	2.36
TOTAL COMPONENT BALANCE		0.60	1.41

Table A-30 Steady State Data for Run FF1 (+20% step in F)

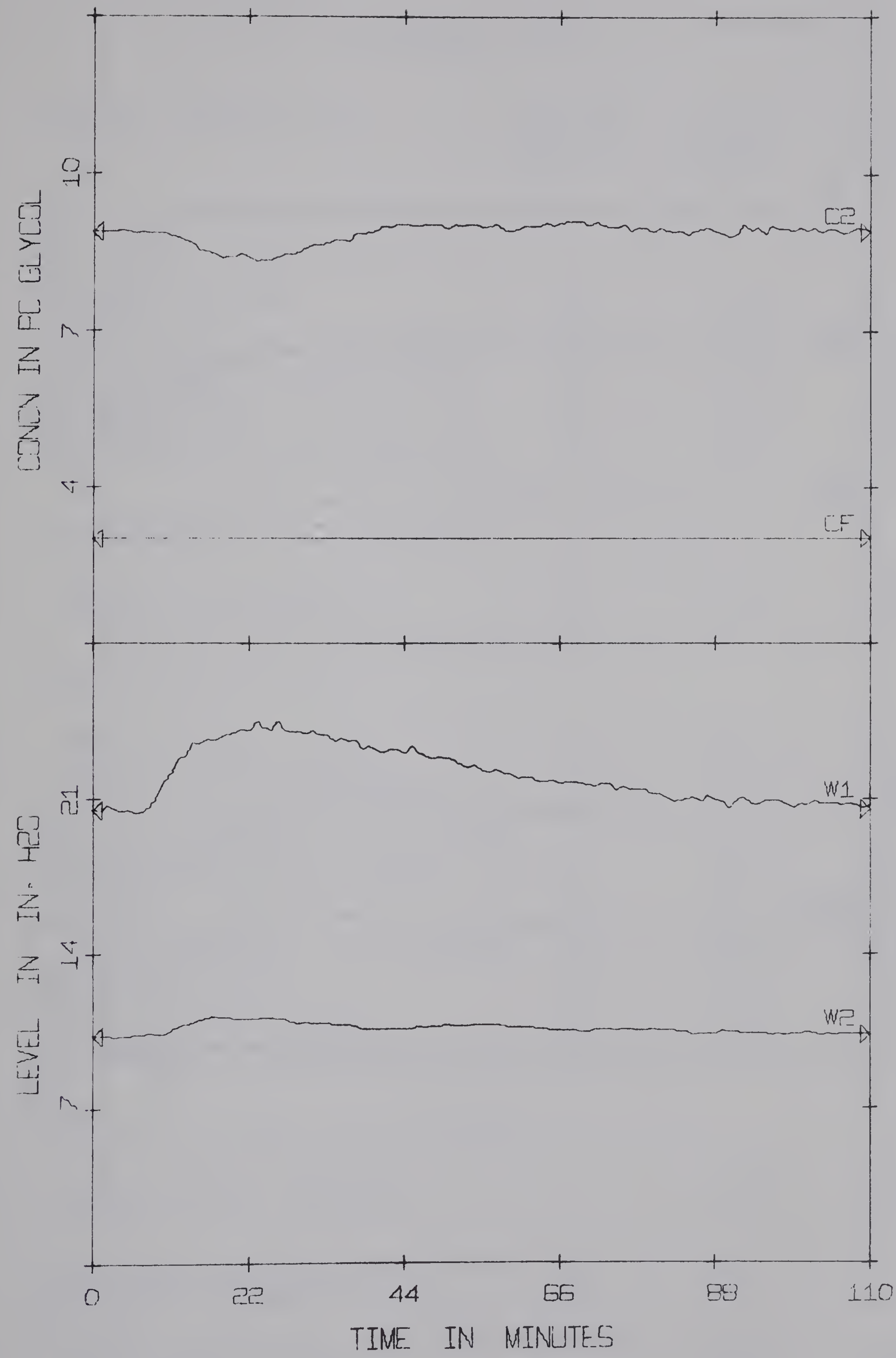


Figure A-29a Transient Data for Run FF1 (+20% step in F)

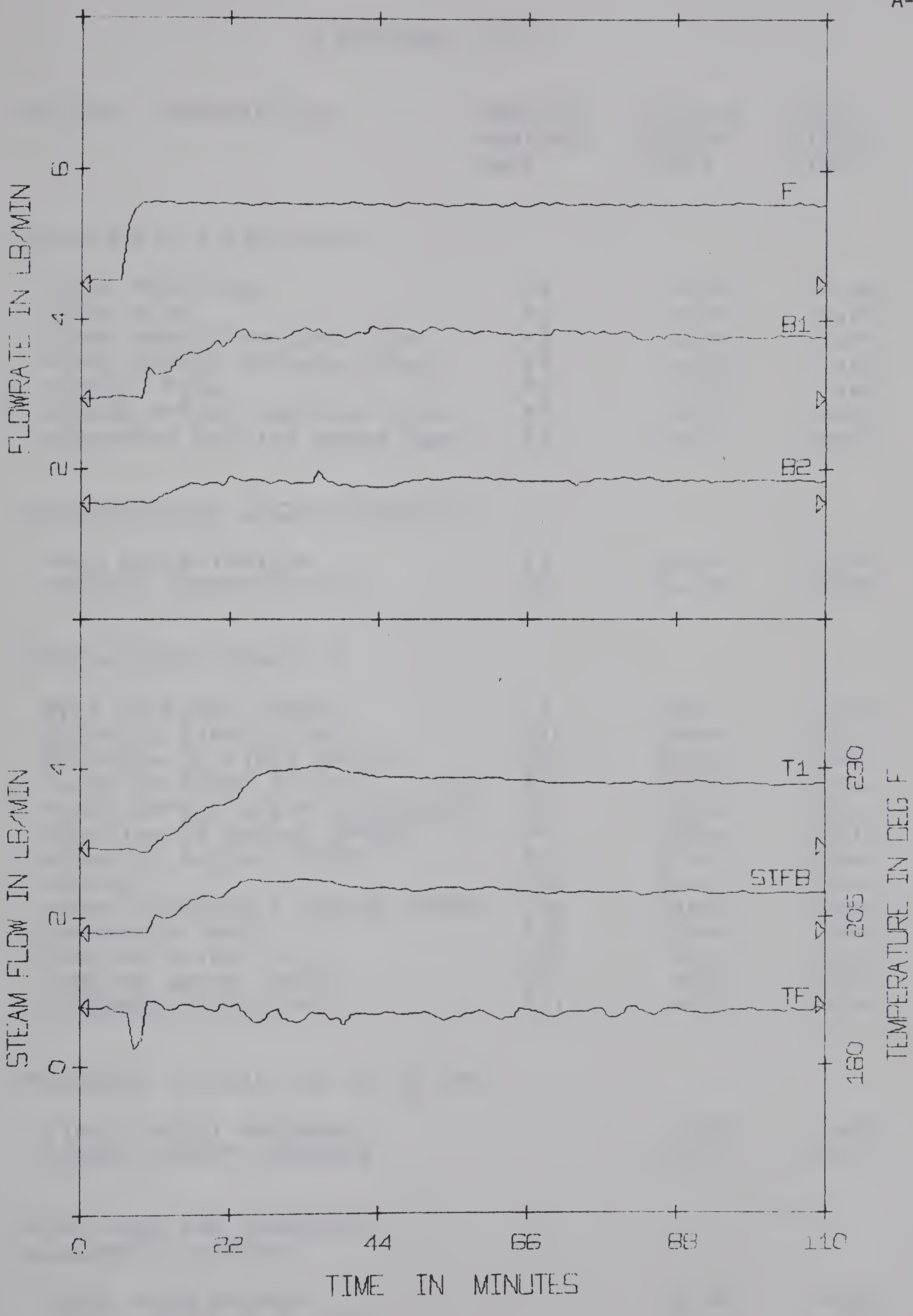
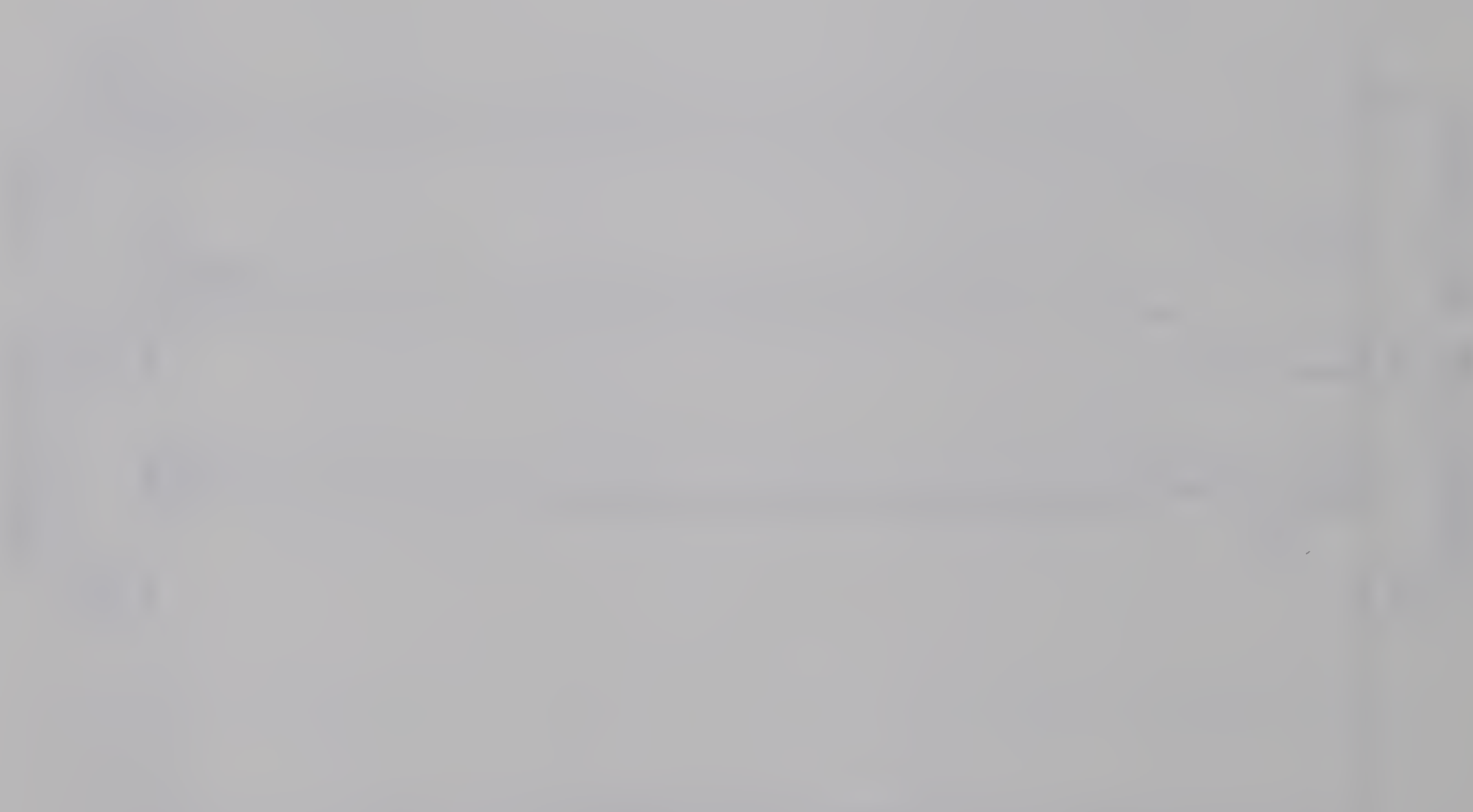


Figure A-29b Transient Data for Run FF1 (+20% step in F)



EXPERIMENT FF2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	5.56
STEAM FLOW	F1	1.79	2.29
FIRST EFFECT BOTTOMS FLOW	F2	2.92	3.74
FIRST EFFECT OVERHEAD FLOW	F5	1.52	1.81
PRODUCT FLOW	F6	1.52	1.80
SECOND EFFECT OVERHEAD FLOW	F7	1.27	1.67
CONDENSER COOLING WATER FLOW	F9	40.11	39.97
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.088	0.088
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.1	190.0
STEAM TO FIRST EFFECT	T15	299.4	297.3
SOLUTION IN FIRST EFFECT	T19	216.0	226.3
VAPOR IN FIRST EFFECT	T2	214.4	224.2
FIRST EFFECT STEAM CONDENSATE	T5	236.2	252.4
SOLUTION TO SECOND EFFECT	T4	180.6	182.2
STEAM TO SECOND EFFECT	T10	214.7	225.0
PRODUCT	T34	155.8	156.6
STEAM CONDENSATE SECOND EFFECT	T28	188.0	196.6
SEPARATOR VAPOR	T12	159.0	158.8
COOLING WATER INLET	T29	52.5	52.0
COOLING WATER OUTLET	T1	90.3	97.7
CONDENSER CONDENSATE	T11	99.3	149.6
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		1.95	7.85
SECOND EFFECT PRESSURE		-13.72	-13.77
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.58	2.98
TOTAL COMPONENT BALANCE		1.10	4.77

Table A-31 Steady State Data for Run FF2 (+20% step in F)

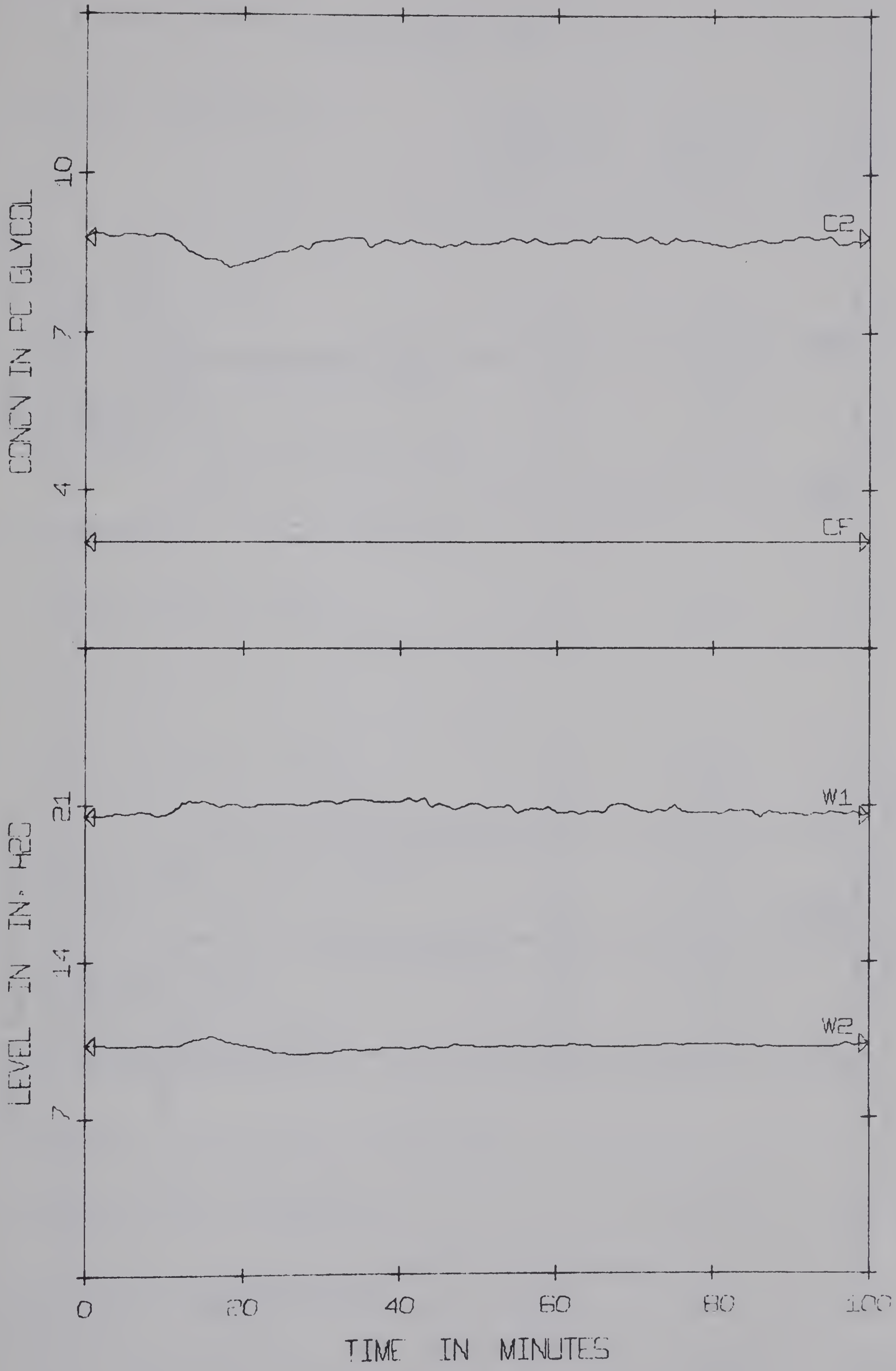


Figure A-30a Transient Data for Run FF2 (+20% step in F)

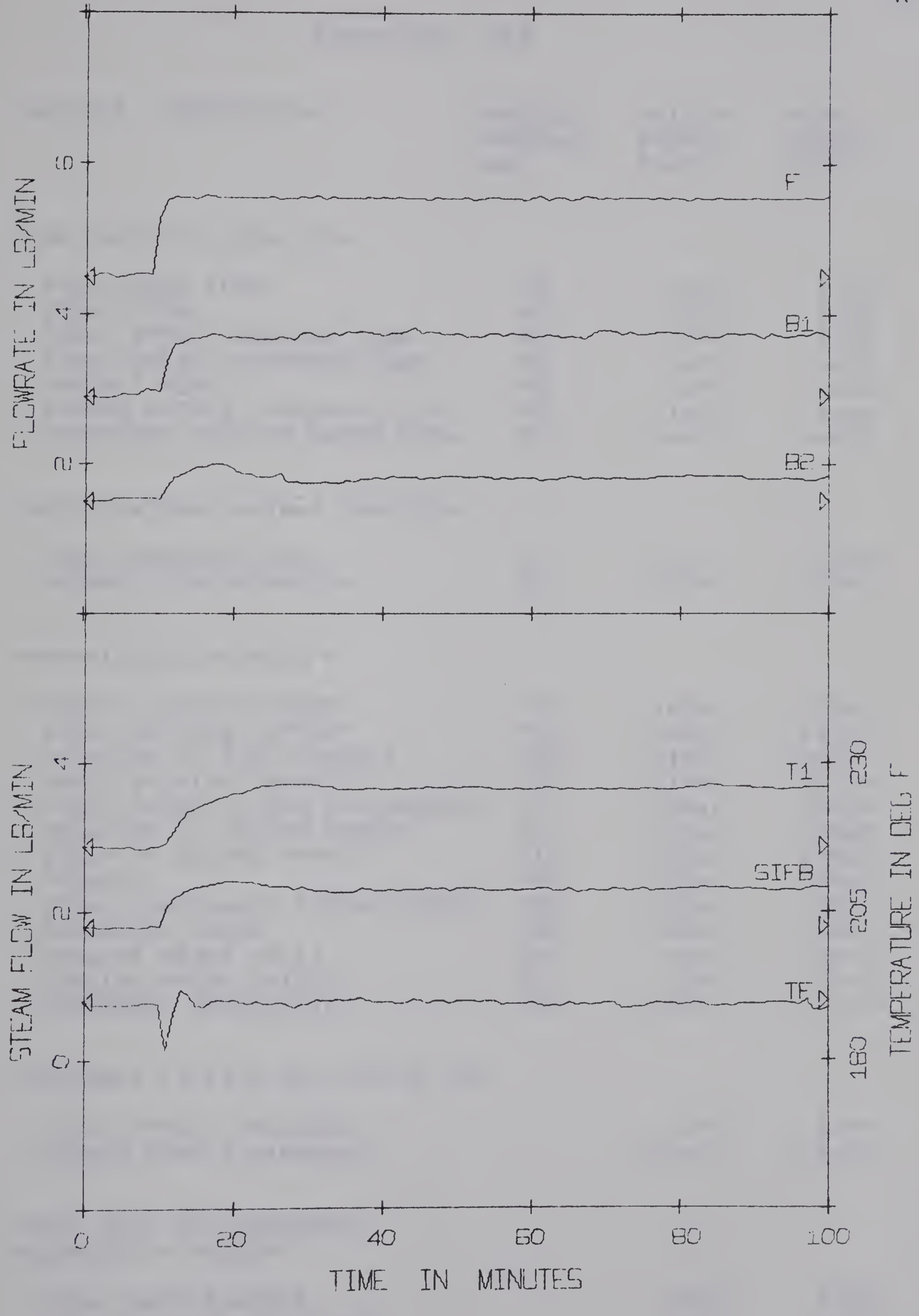


Figure A-30b Transient Data for Run FF2 (+20% step in F)

EXPERIMENT FF3

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	5.56
STEAM FLOW	F1	1.79	2.32
FIRST EFFECT BOTTOMS FLOW	F2	2.88	3.70
FIRST EFFECT OVERHEAD FLOW	F5	1.45	1.79
PRODUCT FLOW	F6	1.48	1.76
SECOND EFFECT OVERHEAD FLOW	F7	1.23	1.66
CONDENSER COOLING WATER FLOW	F9	39.95	40.02
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.031	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.5	189.6
STEAM TO FIRST EFFECT	T15	300.0	297.7
SOLUTION IN FIRST EFFECT	T19	216.7	227.4
VAPOR IN FIRST EFFECT	T2	215.1	225.6
FIRST EFFECT STEAM CONDENSATE	T5	236.8	252.9
SOLUTION TO SECOND EFFECT	T4	181.3	183.2
STEAM TO SECOND EFFECT	T10	215.5	226.0
PRODUCT	T34	157.1	158.6
STEAM CONDENSATE SECOND EFFECT	T28	189.6	198.1
SEPARATOR VAPOR	T12	160.9	160.7
COOLING WATER INLET	T29	54.8	54.6
COOLING WATER OUTLET	T1	92.7	99.7
CONDENSER CONDENSATE	T11	101.7	151.3
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		1.87	8.46
SECOND EFFECT PRESSURE		-13.77	-13.70
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.20	2.63
TOTAL COMPONENT BALANCE		3.36	3.32

Table A-32 Steady State Data for Run FF3 (+20% step in F)

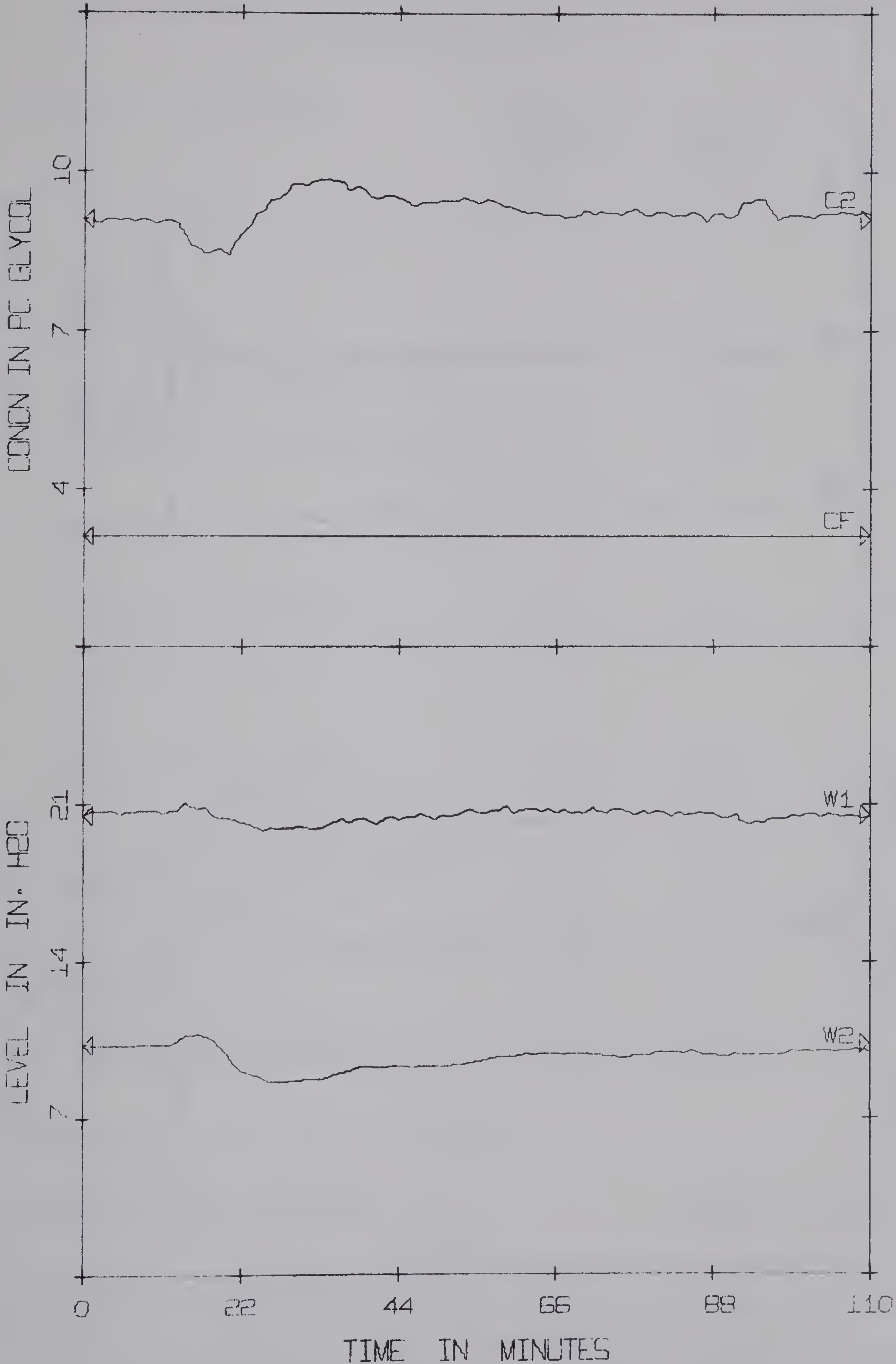


Figure A-31a Transient Data for Run FF3 (+20% step in F)

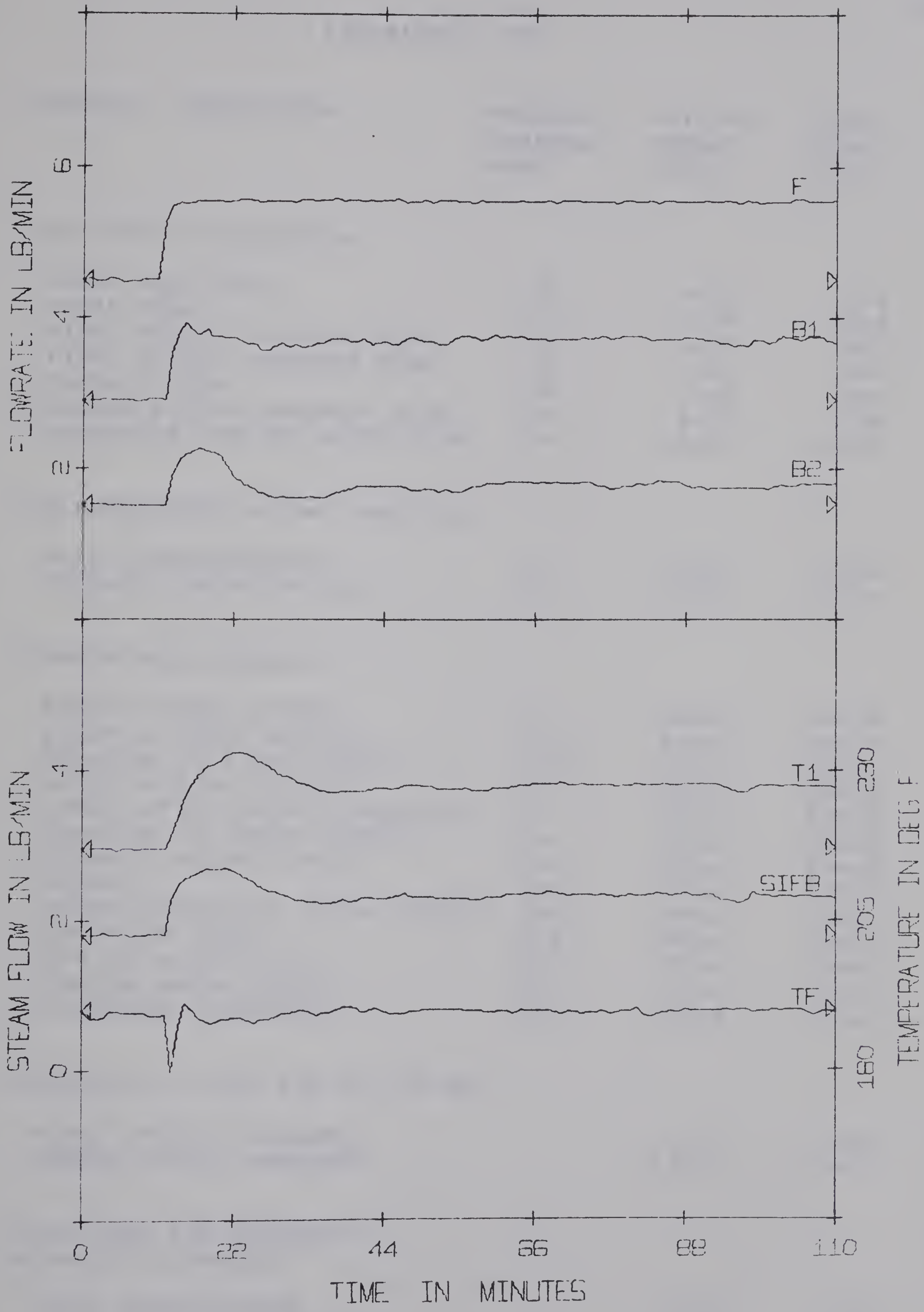


Figure A-31b Transient Data for Run FF3 (+20% step in F)

EXPERIMENT FF5

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.50	5.56
STEAM FLOW	F1	1.79	2.29
FIRST EFFECT BOTTOMS FLOW	F2	2.92	3.85
FIRST EFFECT OVERHEAD FLOW	F5	1.42	1.86
PRODUCT FLOW	F6	1.48	1.81
SECOND EFFECT OVERHEAD FLOW	F7	1.30	1.68
CONDENSER COOLING WATER FLOW	F9	39.87	40.04
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.3	190.6
STEAM TO FIRST EFFECT	T15	300.4	298.4
SOLUTION IN FIRST EFFECT	T19	217.4	226.6
VAPOR IN FIRST EFFECT	T2	215.1	224.3
FIRST EFFECT STEAM CONDENSATE	T5	232.4	249.5
SOLUTION TO SECOND EFFECT	T4	181.8	182.8
STEAM TO SECOND EFFECT	T10	215.4	224.5
PRODUCT	T34	157.3	157.1
STEAM CONDENSATE SECOND EFFECT	T28	189.8	196.3
SEPARATOR VAPOR	T12	160.4	159.3
COOLING WATER INLET	T29	54.4	54.0
COOLING WATER OUTLET	T1	91.1	99.7
CONDENSER CONDENSATE	T11	117.8	151.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		1.77	7.68
SECOND EFFECT PRESSURE		-13.77	-13.75
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.56	3.83
TOTAL COMPONENT BALANCE		1.81	1.67

Table A-33 Steady State Data for Run FF5 (+20% step in F)

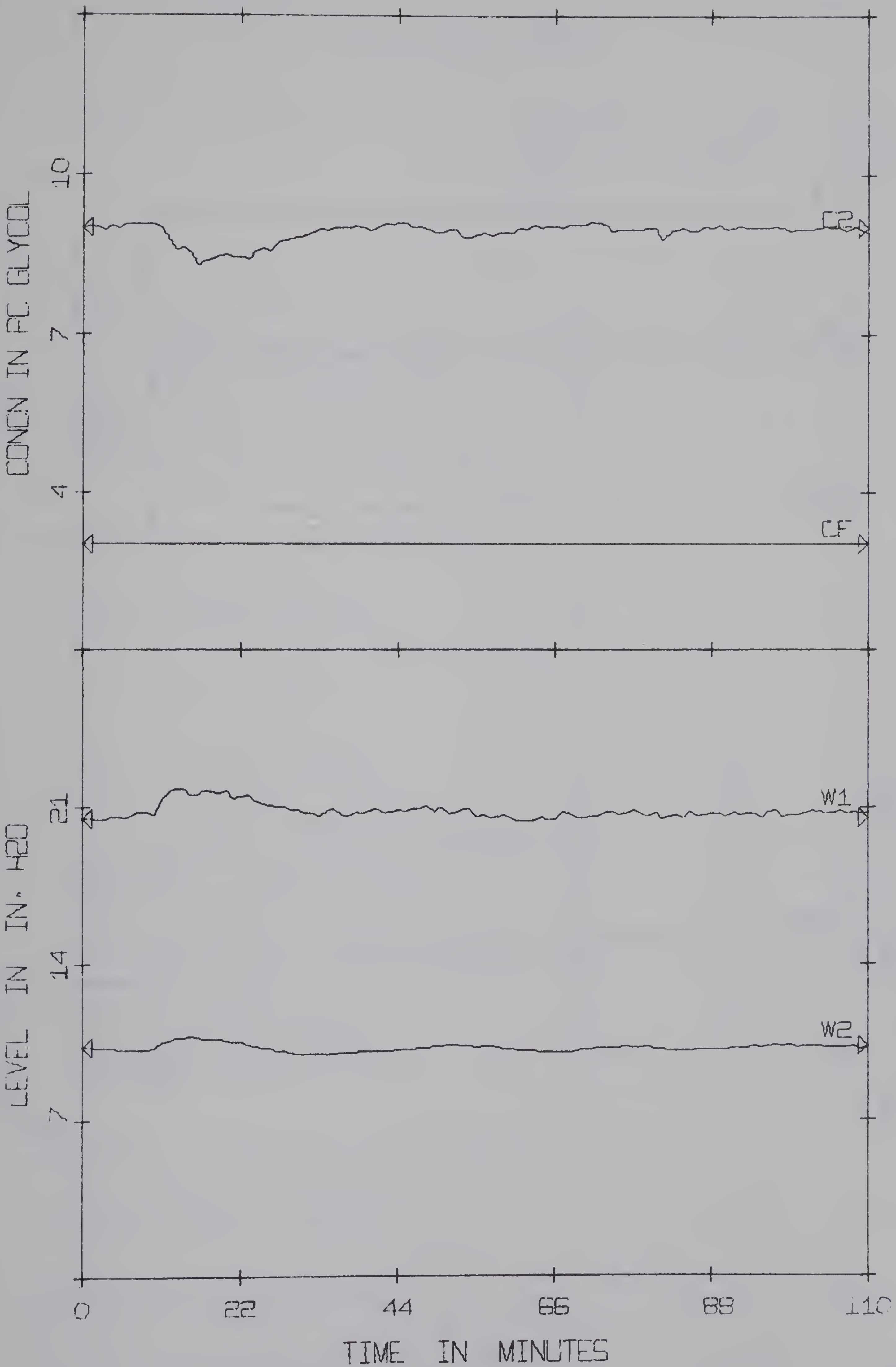


Figure A-32a Transient Data for Run FF5 (+20% step in F)

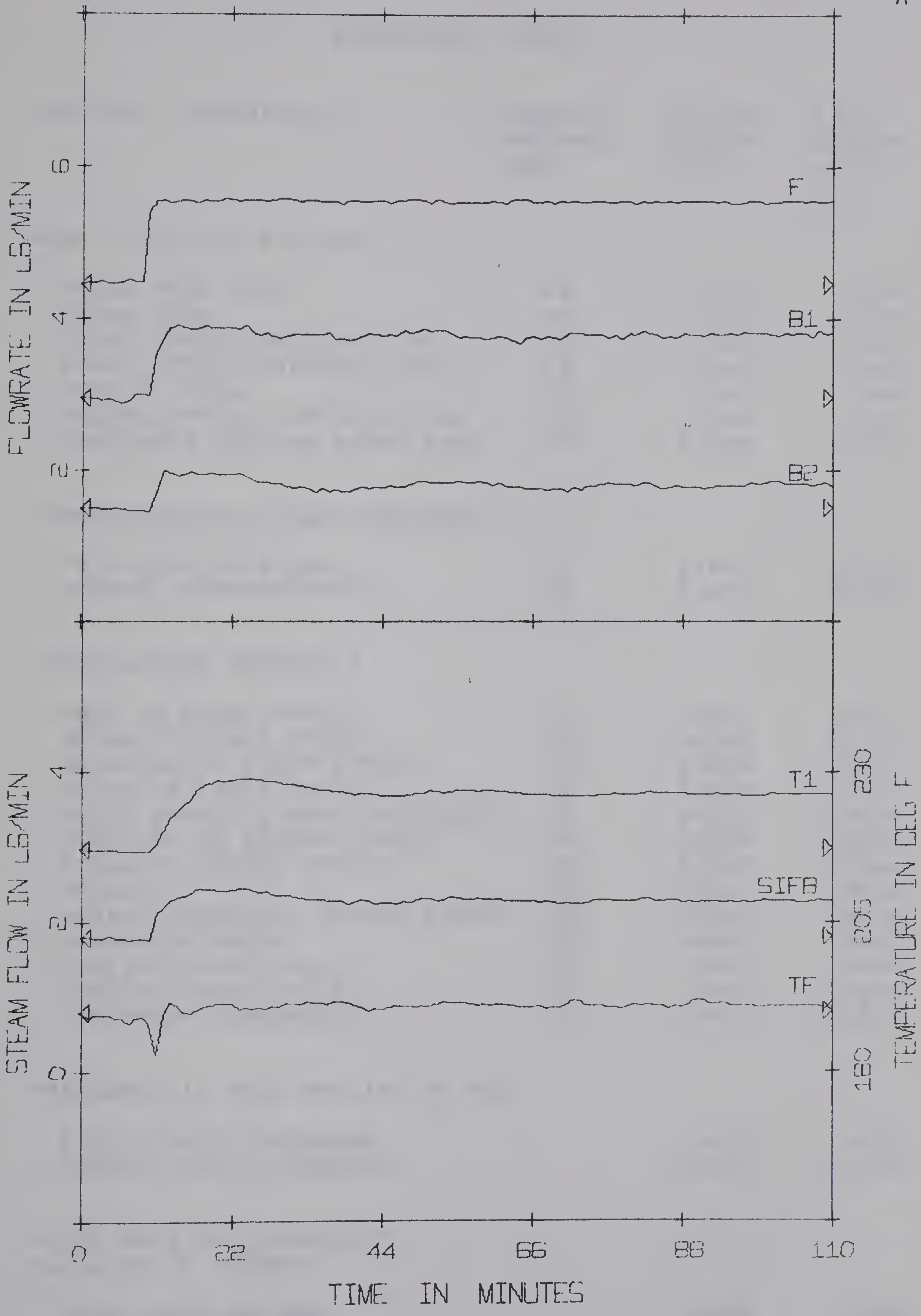


Figure A-32b Transient Data for Run FF5 (+20% step in F)

EXPERIMENT PRED1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.52	5.51
STEAM FLOW	F1	1.81	2.29
FIRST EFFECT BOTTOMS FLOW	F2	2.92	3.61
FIRST EFFECT OVERHEAD FLOW	F5	1.49	1.83
PRODUCT FLOW	F6	1.81	1.86
SECOND EFFECT OVERHEAD FLOW	F7	1.32	1.72
CONDENSER COOLING WATER FLOW	F9	39.69	39.91
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.2	188.0
STEAM TO FIRST EFFECT	T15	300.5	297.7
SOLUTION IN FIRST EFFECT	T19	222.6	255.7
VAPOR IN FIRST EFFECT	T2	220.9	223.9
FIRST EFFECT STEAM CONDENSATE	T5	242.1	249.9
SOLUTION TO SECOND EFFECT	T4	180.8	182.0
STEAM TO SECOND EFFECT	T10	221.3	224.2
PRODUCT	T34	157.4	158.6
STEAM CONDENSATE SECOND EFFECT	T28	199.6	198.9
SEPARATOR VAPOR	T12	160.2	160.2
COOLING WATER INLET	T29	49.4	48.5
COOLING WATER OUTLET	T1	89.0	93.5
CONDENSER CONDENSATE	T11	130.0	143.7
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.25	8.13
SECOND EFFECT PRESSURE		-15.85	-15.88
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.24	2.48
TOTAL COMPONENT BALANCE		-0.60	-0.81

Table A-34 Steady State Data for Run PRED1 (+20% step in F)

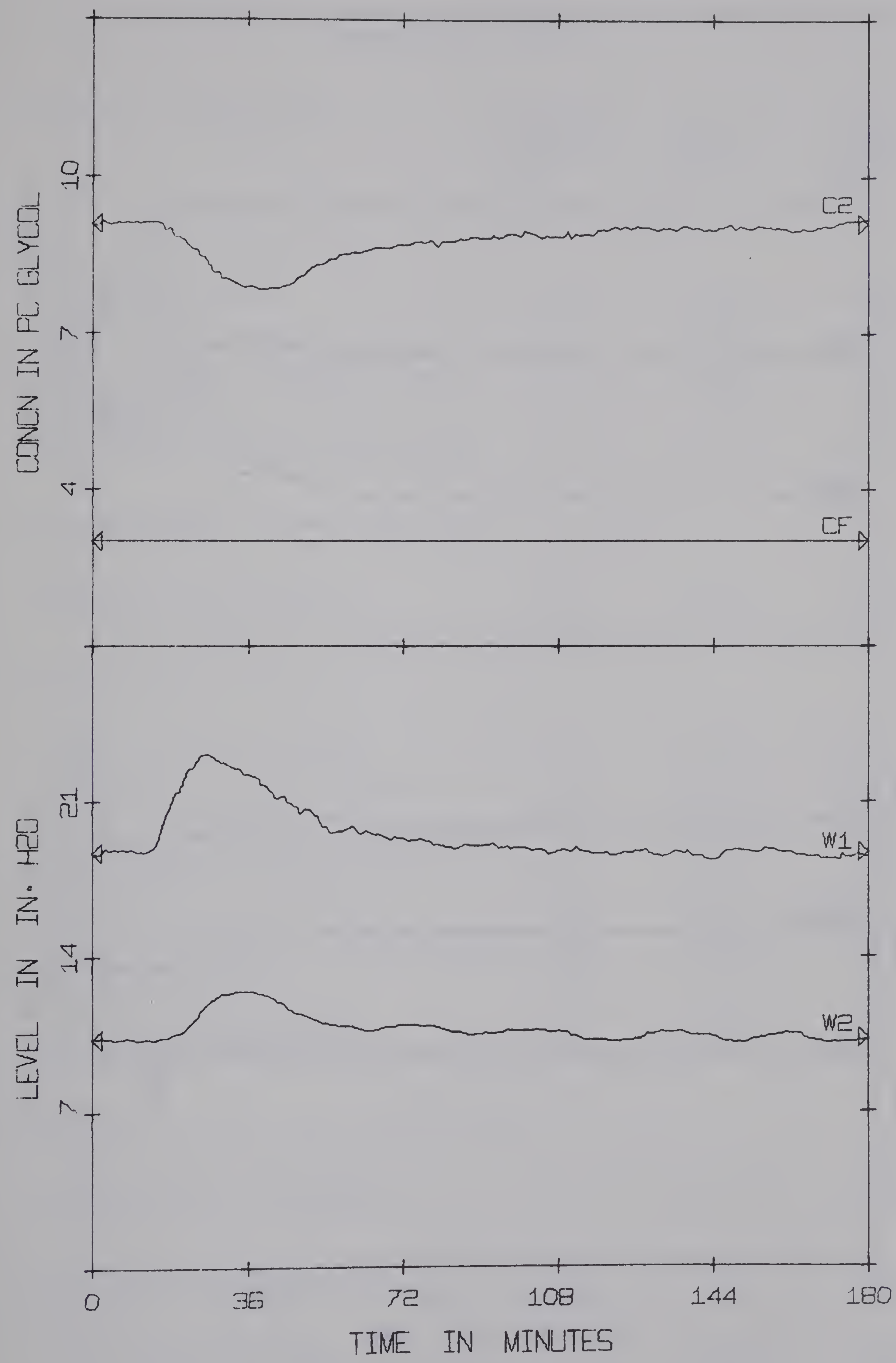


Figure A-33a Transient Data for Run PRED1 (+20% step in F)

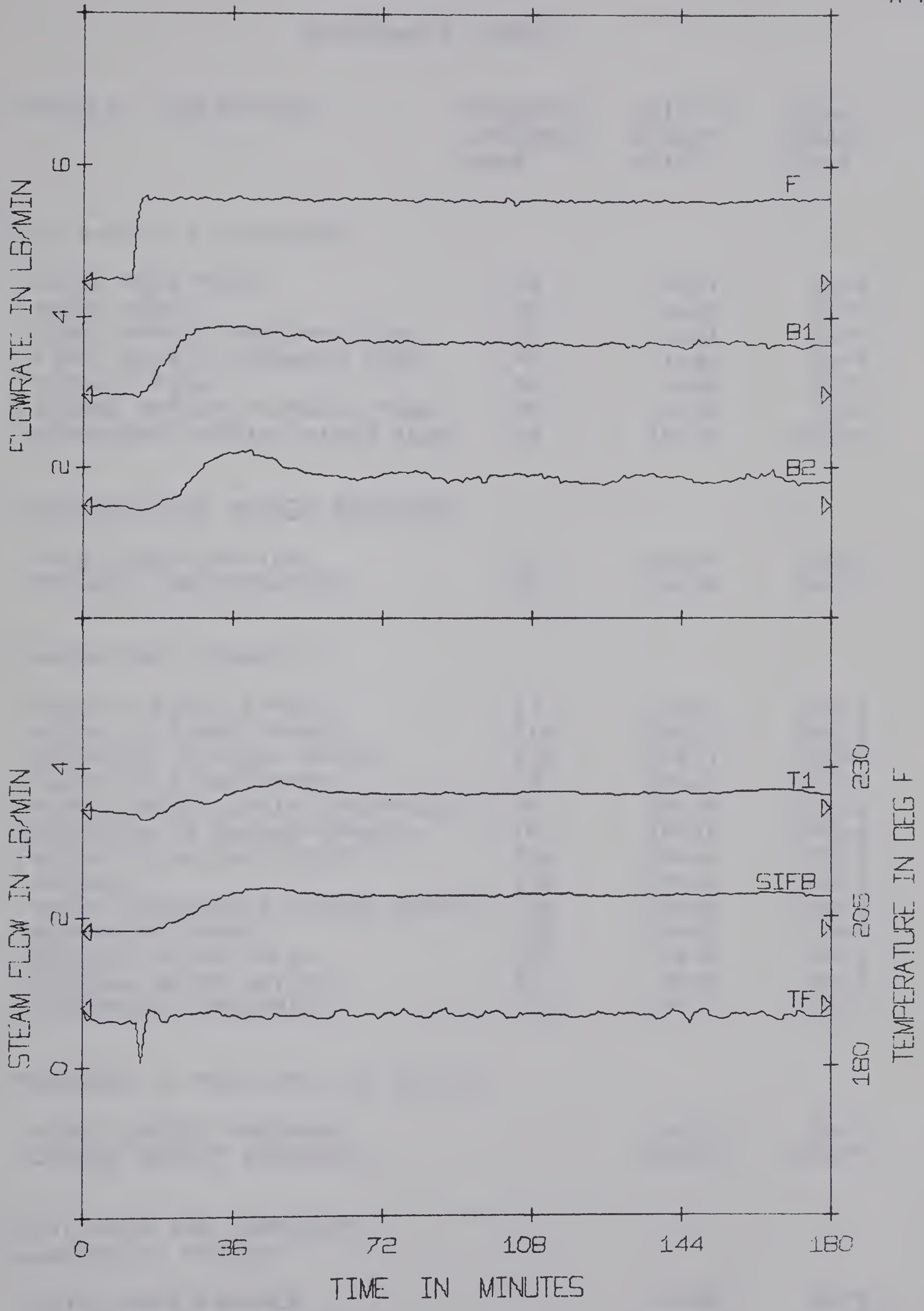


Figure A-33b Transient Data for Run PRED1 (+20% step in F)

EXPERIMENT PRED2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.51	4.48
STEAM FLOW	F1	2.29	1.77
FIRST EFFECT BOTTOMS FLOW	F2	3.61	2.97
FIRST EFFECT OVERHEAD FLOW	F5	1.83	1.48
PRODUCT FLOW	F6	1.86	1.77
SECOND EFFECT OVERHEAD FLOW	F7	1.72	1.31
CONDENSER COOLING WATER FLOW	F9	39.91	40.01
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.0	188.3
STEAM TO FIRST EFFECT	T15	297.7	301.3
SOLUTION IN FIRST EFFECT	T19	255.7	222.5
VAPOR IN FIRST EFFECT	T2	223.9	220.9
FIRST EFFECT STEAM CONDENSATE	T5	249.9	242.0
SOLUTION TO SECOND EFFECT	T4	182.0	180.9
STEAM TO SECOND EFFECT	T10	224.2	221.3
PRODUCT	T34	158.6	157.3
STEAM CONDENSATE SECOND EFFECT	T28	198.9	194.5
SEPARATOR VAPOR	T12	160.2	159.8
COOLING WATER INLET	T29	48.5	48.1
COOLING WATER OUTLET	T1	93.5	85.9
CONDENSER CONDENSATE	T11	143.7	110.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		8.13	6.07
SECOND EFFECT PRESSURE		-15.88	-15.89
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.48	3.75
TOTAL COMPONENT BALANCE		-0.81	-1.30

Table A-35 Steady State Data for Run PRED2 (-20% step in F)

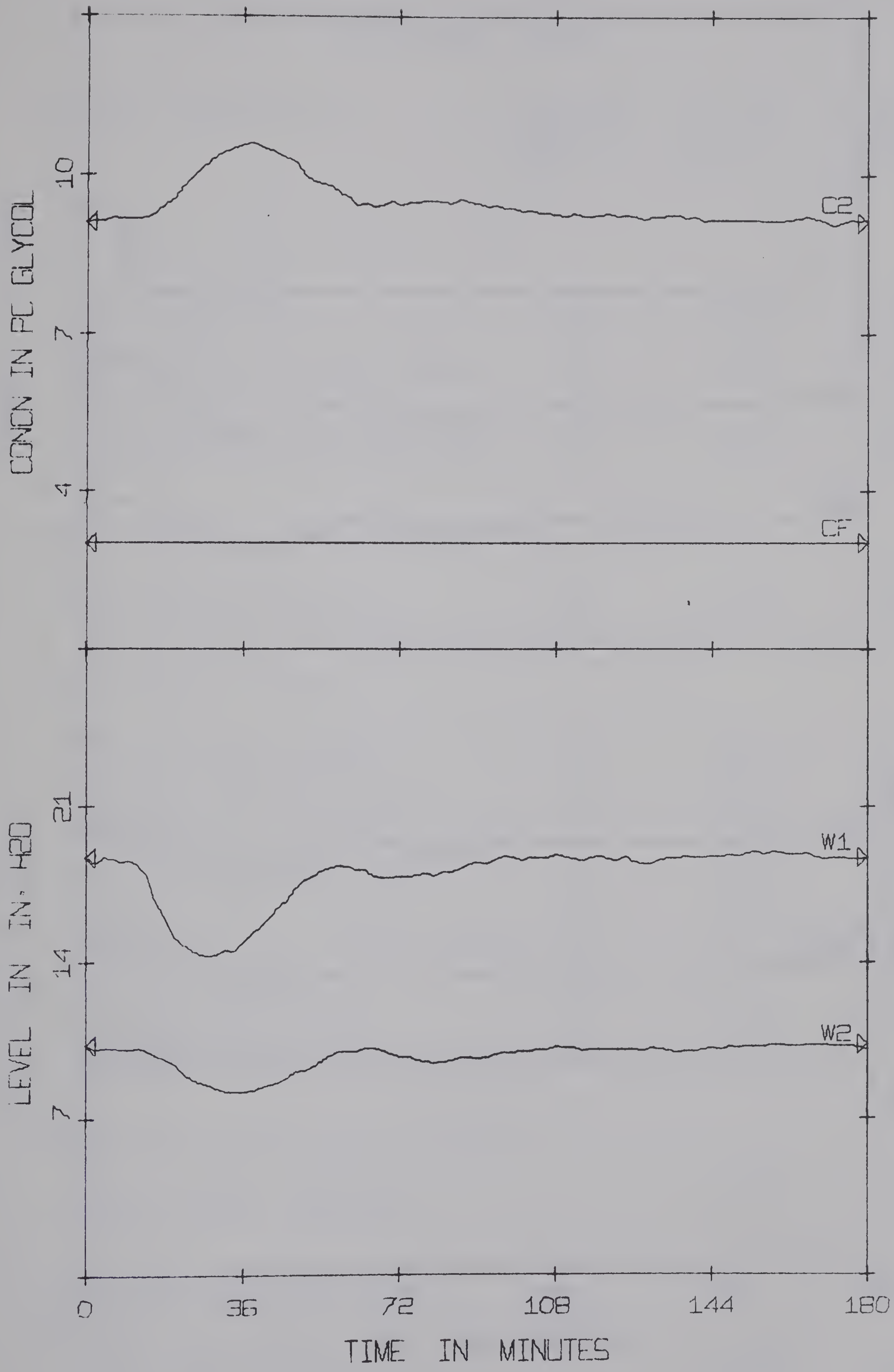


Figure A-34a Transient Data for Run PRED2 (-20% step in F)

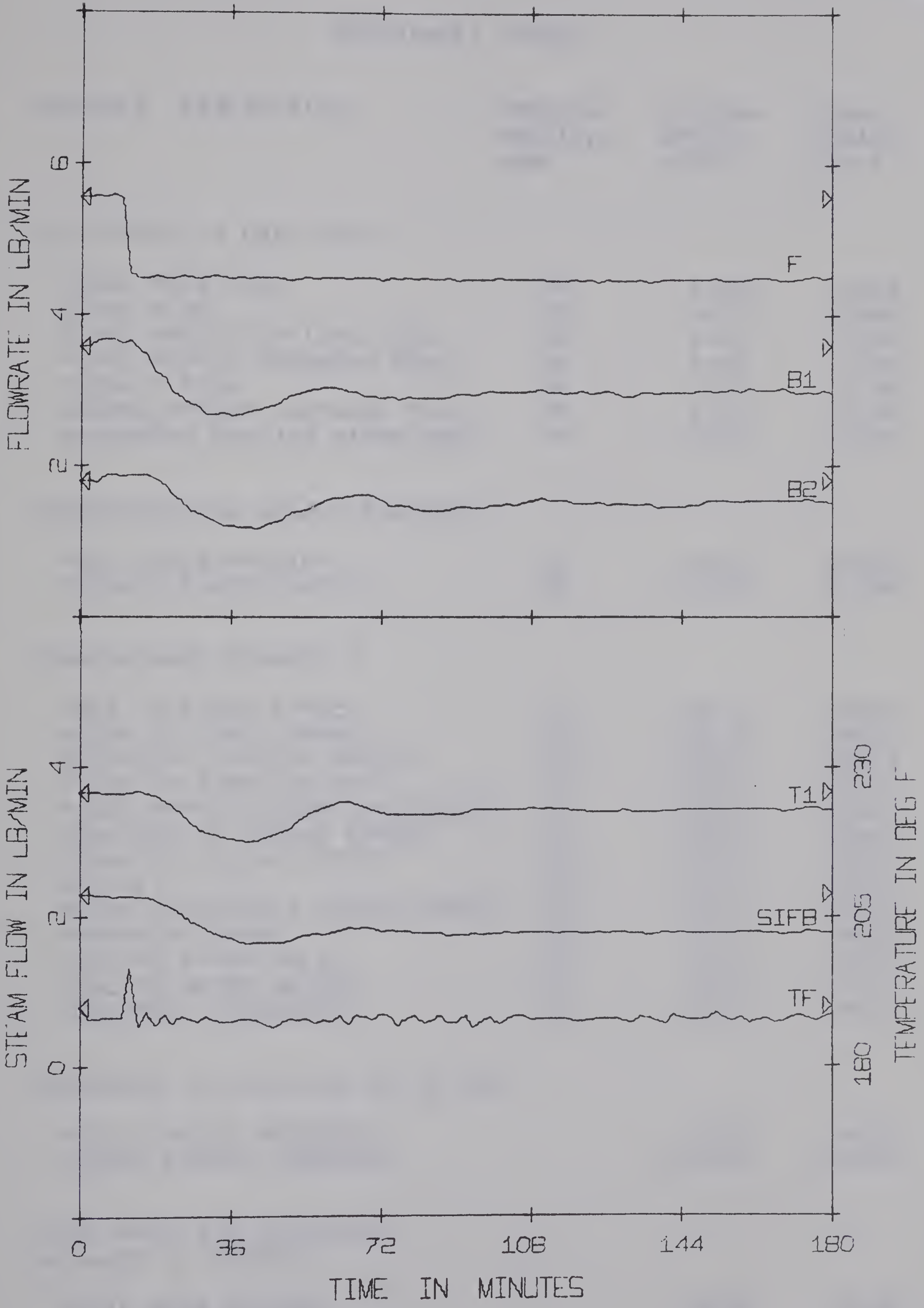


Figure A-34b Transient Data for Run PRED2 (-20% step in F)

EXPERIMENT PRED3

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	5.55
STEAM FLOW	F1	1.77	2.30
FIRST EFFECT BOTTOMS FLOW	F2	2.97	3.64
FIRST EFFECT OVERHEAD FLOW	F5	1.48	1.83
PRODUCT FLOW	F6	1.77	1.79
SECOND EFFECT OVERHEAD FLOW	F7	1.31	1.68
CONDENSER COOLING WATER FLOW	F9	40.01	40.09
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.3	188.2
STEAM TO FIRST EFFECT	T15	301.3	297.6
SOLUTION IN FIRST EFFECT	T19	222.5	226.3
VAPOR IN FIRST EFFECT	T2	220.9	224.5
FIRST EFFECT STEAM CONDENSATE	T5	242.0	250.5
SOLUTION TO SECOND EFFECT	T4	180.9	183.3
STEAM TO SECOND EFFECT	T10	221.3	225.0
PRODUCT	T34	157.3	158.7
STEAM CONDENSATE SECOND EFFECT	T28	194.5	200.0
SEPARATOR VAPOR	T12	159.8	160.8
COOLING WATER INLET	T29	48.1	51.9
COOLING WATER OUTLET	T1	85.9	97.1
CONDENSER CONDENSATE	T11	110.0	149.5
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		6.07	8.27
SECOND EFFECT PRESSURE		-15.89	-15.82
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.75	2.76
TOTAL COMPONENT BALANCE		-1.30	2.83

Table A-36 Steady State Data for Run PRED3 (+20% step in F)

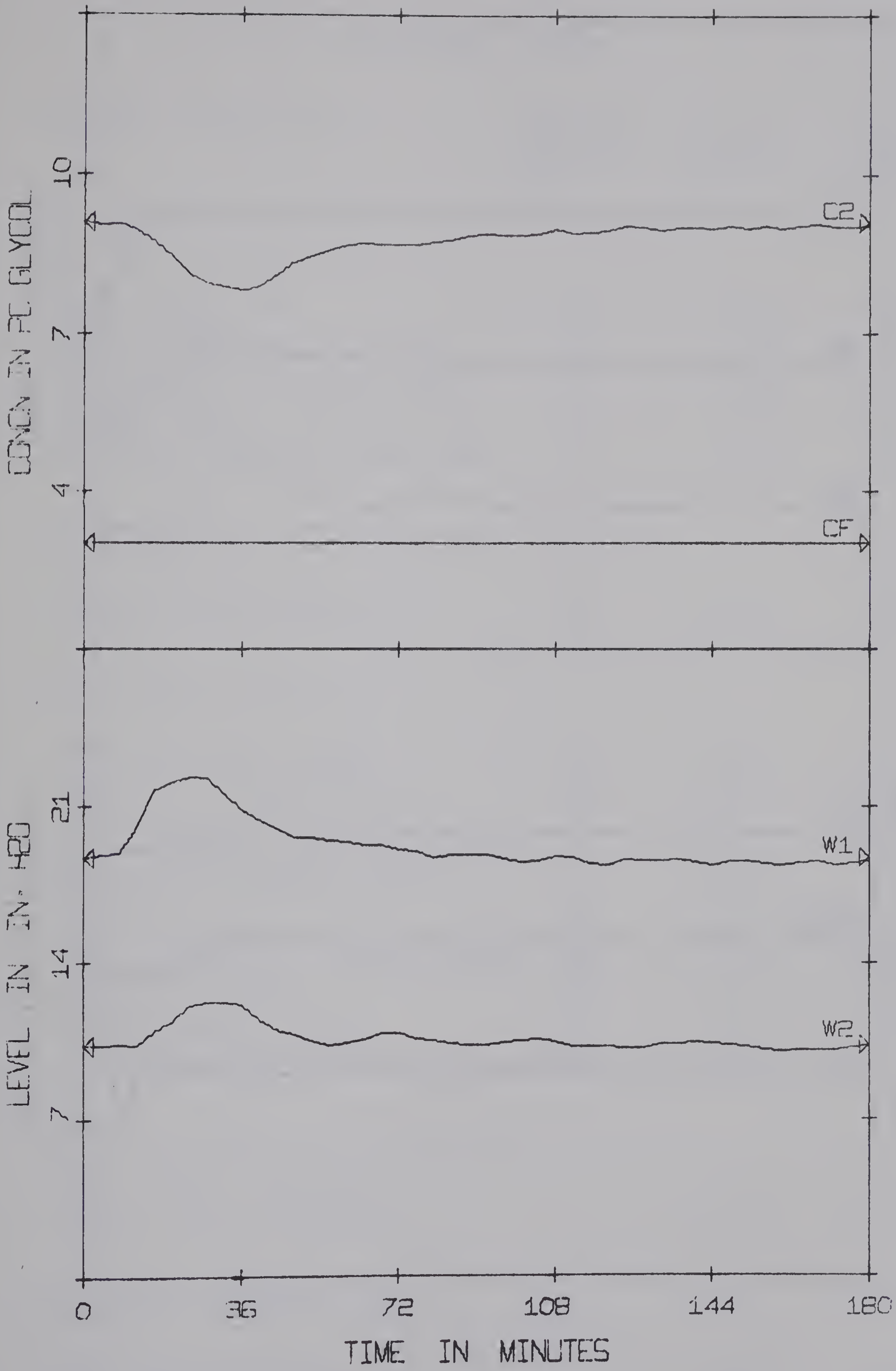
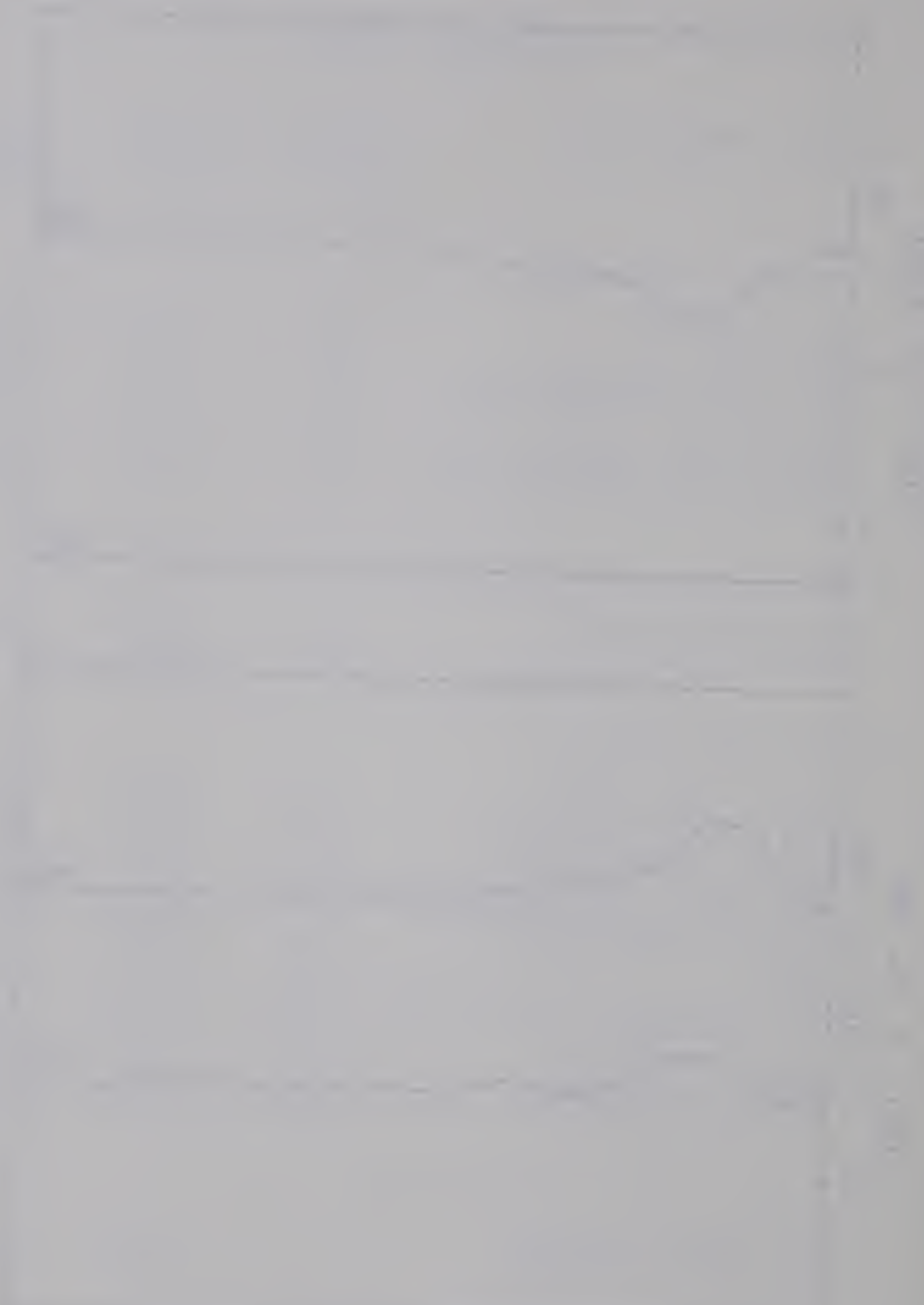


Figure A-35a Transient Data for Run PRED3 (+20% step in F)



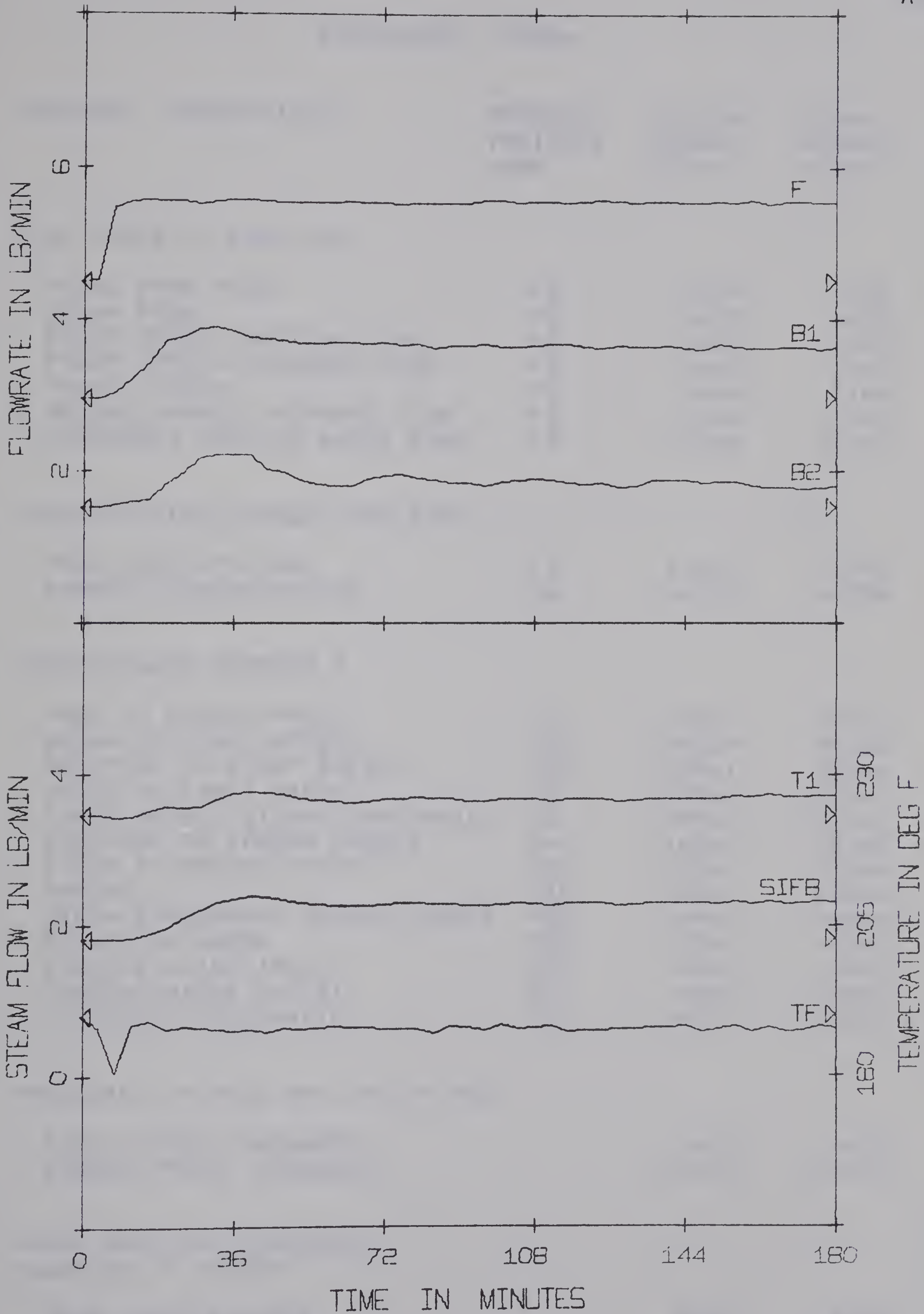


Figure A-35b Transient Data for Run PRED3 (+20% step in F)

EXPERIMENT PRED4

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.54	4.45
STEAM FLOW	F1	2.28	1.79
FIRST EFFECT BOTTOMS FLOW	F2	3.60	2.97
FIRST EFFECT OVERHEAD FLOW	F5	1.82	1.47
PRODUCT FLOW	F6	1.84	1.48
SECOND EFFECT OVERHEAD FLOW	F7	1.69	1.27
CONDENSER COOLING WATER FLOW	F9	40.08	39.81
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	187.7	188.6
STEAM TO FIRST EFFECT	T15	297.8	301.3
SOLUTION IN FIRST EFFECT	T19	225.1	222.6
VAPOR IN FIRST EFFECT	T2	223.4	221.0
FIRST EFFECT STEAM CONDENSATE	T5	250.0	242.4
SOLUTION TO SECOND EFFECT	T4	181.7	181.8
STEAM TO SECOND EFFECT	T10	223.7	221.4
PRODUCT	T34	158.3	158.3
STEAM CONDENSATE SECOND EFFECT	T28	199.0	194.8
SEPARATOR VAPOR	T12	159.9	161.1
COOLING WATER INLET	T29	48.1	48.7
COOLING WATER OUTLET	T1	92.0	88.0
CONDENSER CONDENSATE	T11	143.1	120.0
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		7.93	6.29
SECOND EFFECT PRESSURE		-15.83	-15.83
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.17	3.03
TOTAL COMPONENT BALANCE		-0.50	-1.50

Table A-37 Steady State Data for Run PRED4 (-20% step in F)

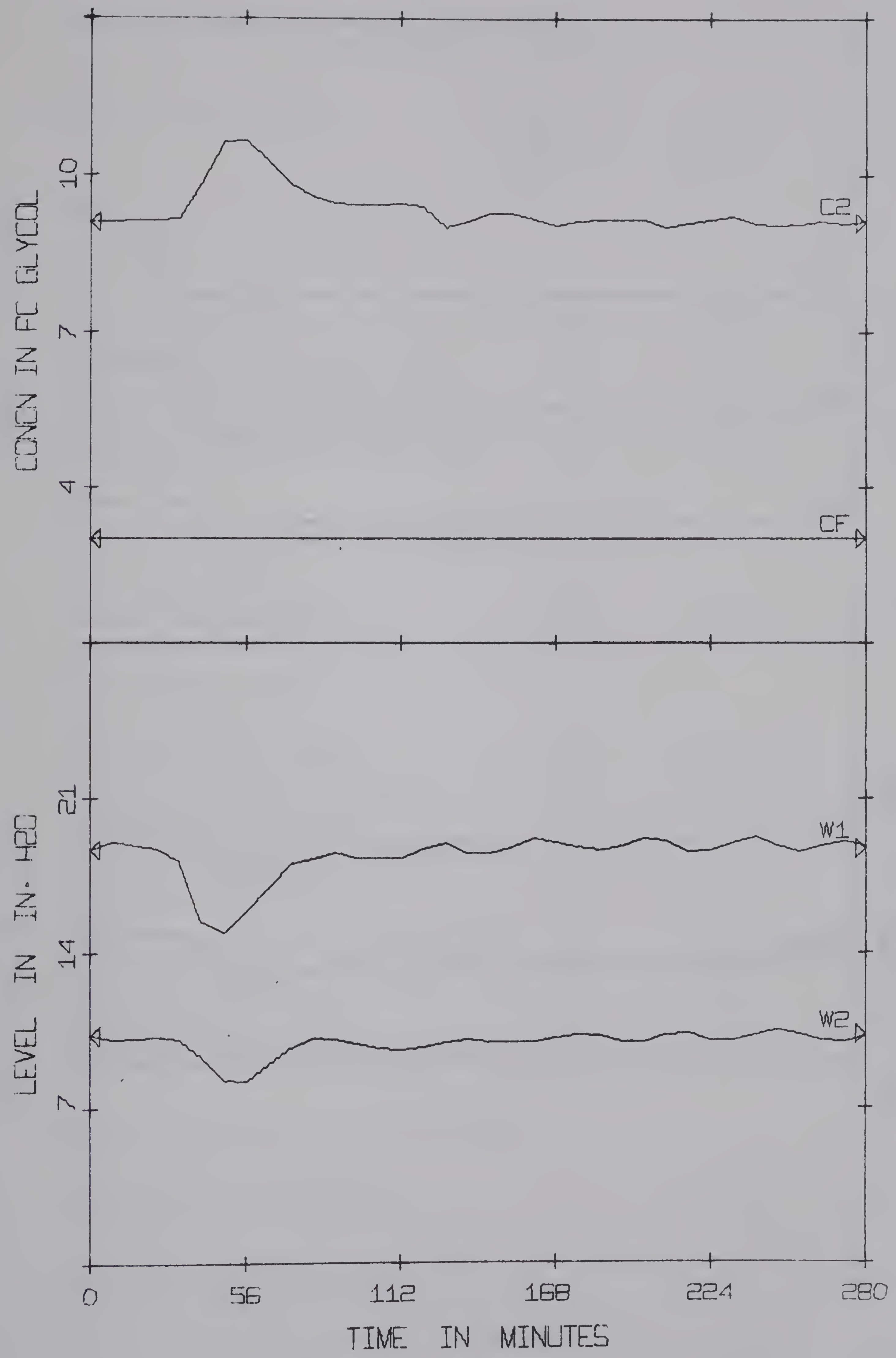


Figure A-36a Transient Data for Run PRED4 (-20% step in F)

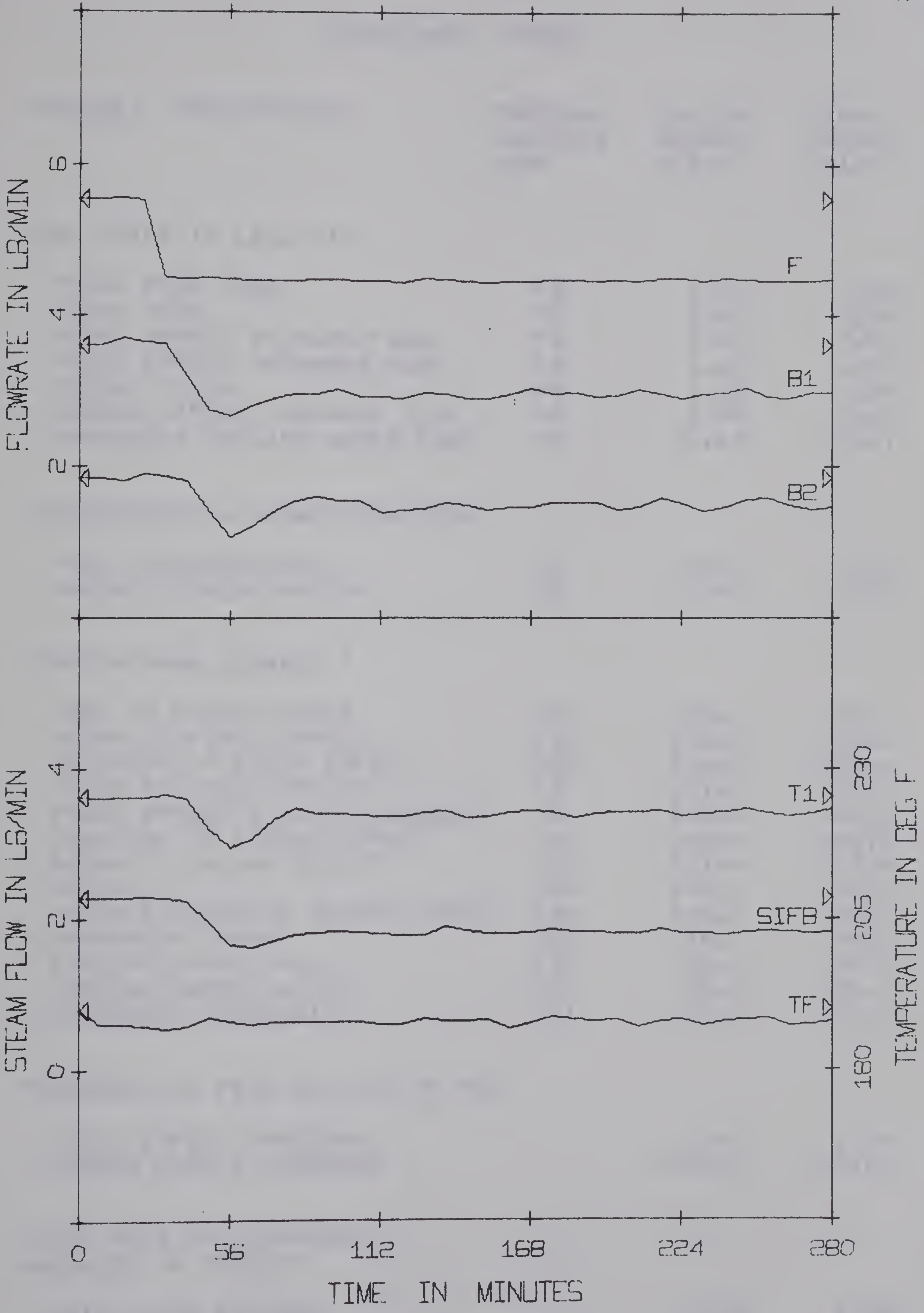


Figure A-36b Transient Data for Run PRED4 (-20% step in F)

EXPERIMENT PRED5

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.01	4.96
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	2.97	2.91
FIRST EFFECT OVERHEAD FLOW	F5	1.63	1.53
PRODUCT FLOW	F6	1.66	1.64
SECOND EFFECT OVERHEAD FLOW	F7	1.55	1.52
CONDENSER COOLING WATER FLOW	F9	39.83	39.81
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.1	190.1
STEAM TO FIRST EFFECT	T15	300.5	300.0
SOLUTION IN FIRST EFFECT	T19	219.2	219.4
VAPOR IN FIRST EFFECT	T2	217.2	217.2
FIRST EFFECT STEAM CONDENSATE	T5	240.6	241.0
SOLUTION TO SECOND EFFECT	T4	181.8	181.2
STEAM TO SECOND EFFECT	T10	217.6	217.8
PRODUCT	T34	155.7	155.9
STEAM CONDENSATE SECOND EFFECT	T28	191.9	195.0
SEPARATOR VAPOR	T12	158.2	158.6
COOLING WATER INLET	T29	52.1	55.3
COOLING WATER OUTLET	T1	92.6	96.5
CONDENSER CONDENSATE	T11	137.3	145.5
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.45	2.48
SECOND EFFECT PRESSURE		-12.01	-12.01
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.14	2.36
TOTAL COMPONENT BALANCE		0.88	-0.58

Table A-38 Steady State Data for Run PRED5 (Change C2 setpoint)

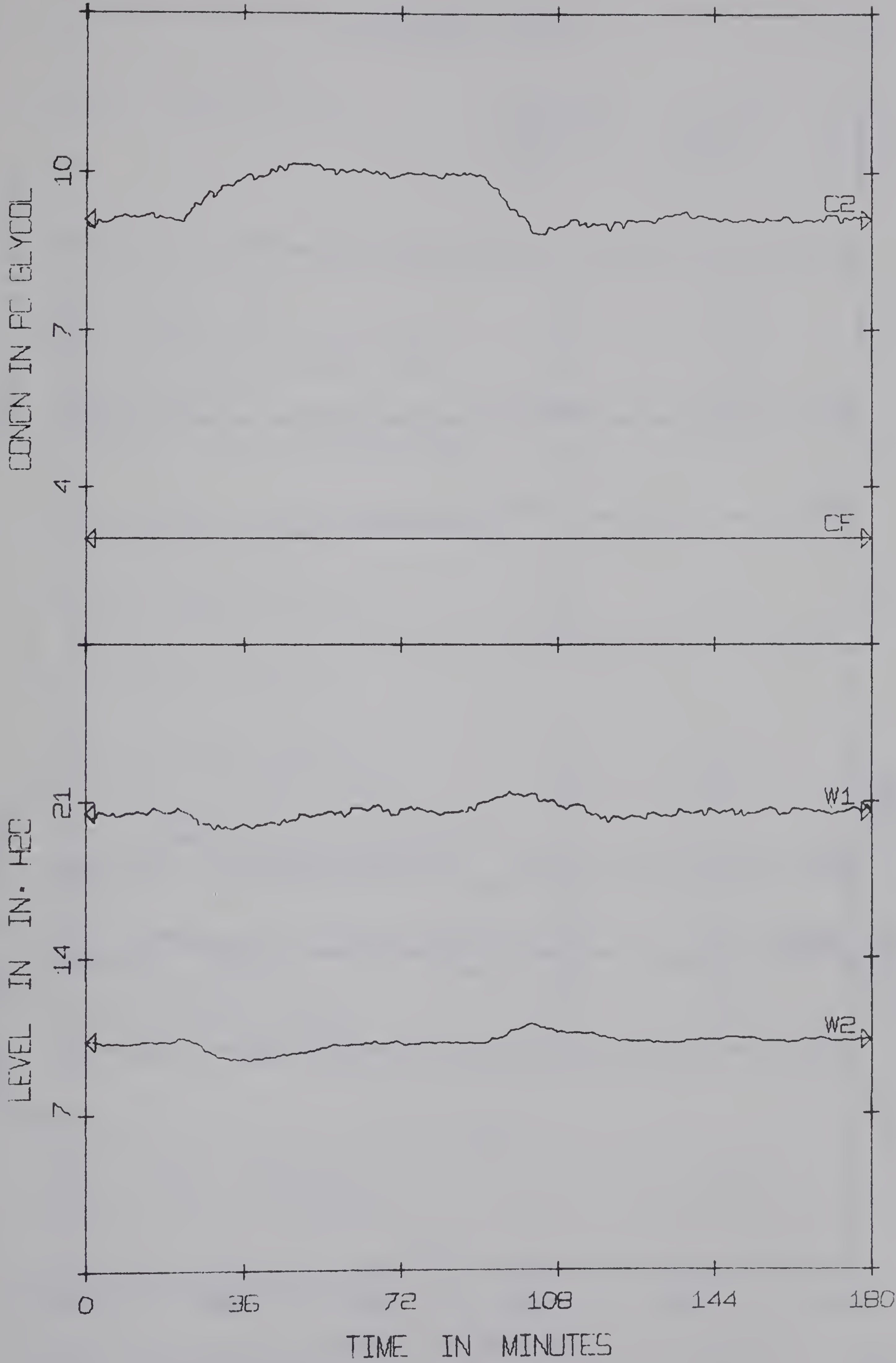


Figure A-37a Transient Data for Run PRED5 (Change C2 setpoint)

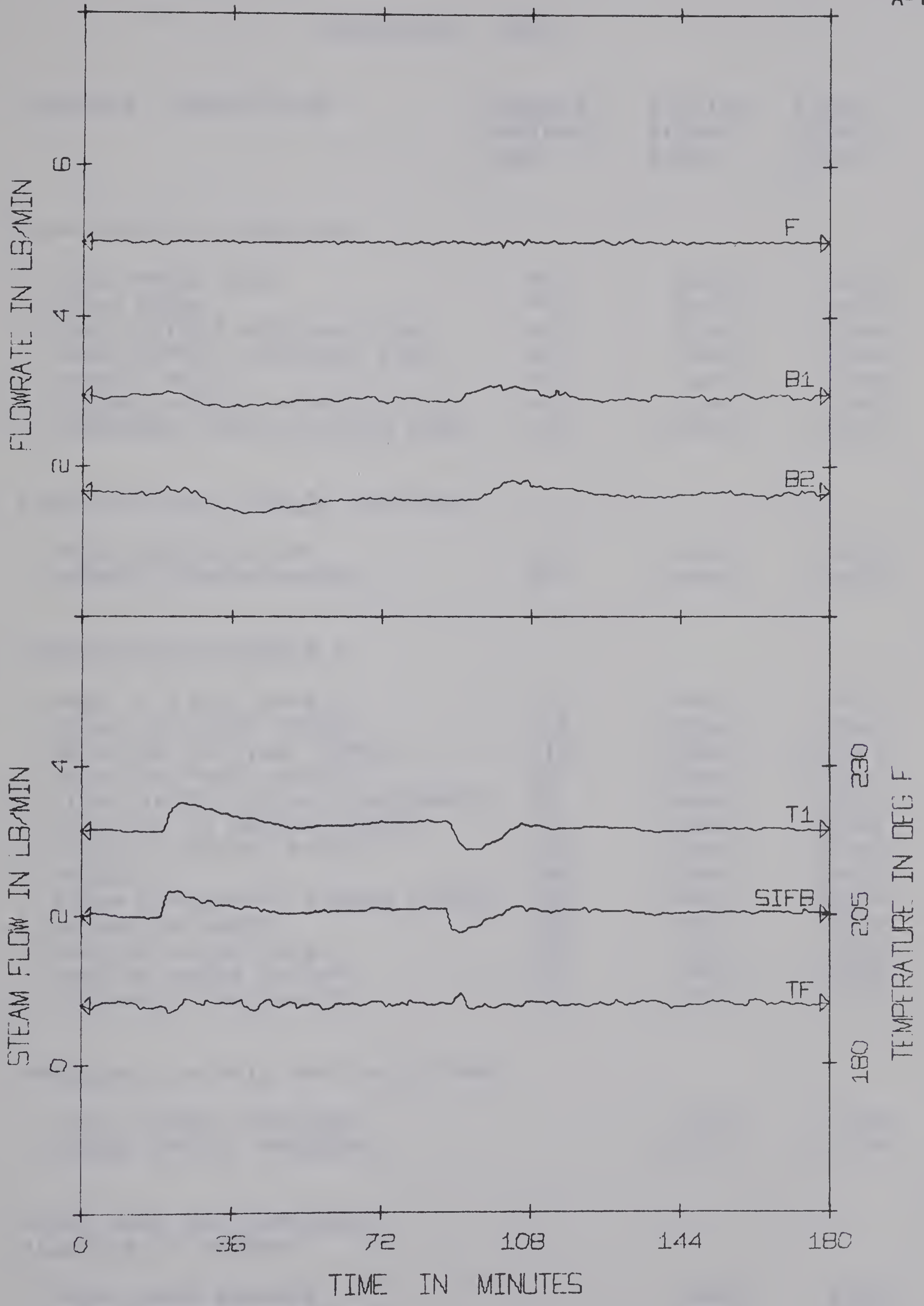


Figure A-37b Transient Data for Run PRED5 (Change C2 setpoint)

EXPERIMENT LSU1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.00	4.94
STEAM FLOW	F1	2.00	2.00
FIRST EFFECT BOTTOMS FLOW	F2	2.92	2.94
FIRST EFFECT OVERHEAD FLOW	F5	1.66	1.65
PRODUCT FLOW	F6	1.61	1.64
SECOND EFFECT OVERHEAD FLOW	F7	1.51	1.52
CONDENSER COOLING WATER FLOW	F9	40.00	40.07
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.093	0.093
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.6	189.7
STEAM TO FIRST EFFECT	T15	301.0	300.8
SOLUTION IN FIRST EFFECT	T19	218.3	218.2
VAPOR IN FIRST EFFECT	T2	216.4	216.1
FIRST EFFECT STEAM CONDENSATE	T5	240.2	239.7
SOLUTION TO SECOND EFFECT	T4	180.3	180.8
STEAM TO SECOND EFFECT	T10	216.8	216.5
PRODUCT	T34	154.4	154.7
STEAM CONDENSATE SECOND EFFECT	T28	192.5	193.4
SEPARATOR VAPOR	T12	156.7	157.5
COOLING WATER INLET	T29	53.3	53.0
COOLING WATER OUTLET	T1	94.1	93.8
CONDENSER CONDENSATE	T11	127.4	127.2
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		1.91	1.88
SECOND EFFECT PRESSURE		-11.97	-11.98
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.32	2.63
TOTAL COMPONENT BALANCE		0.45	-0.77

Table A-39 Steady State Data for Run LSU1 (Change C2 setpoint)

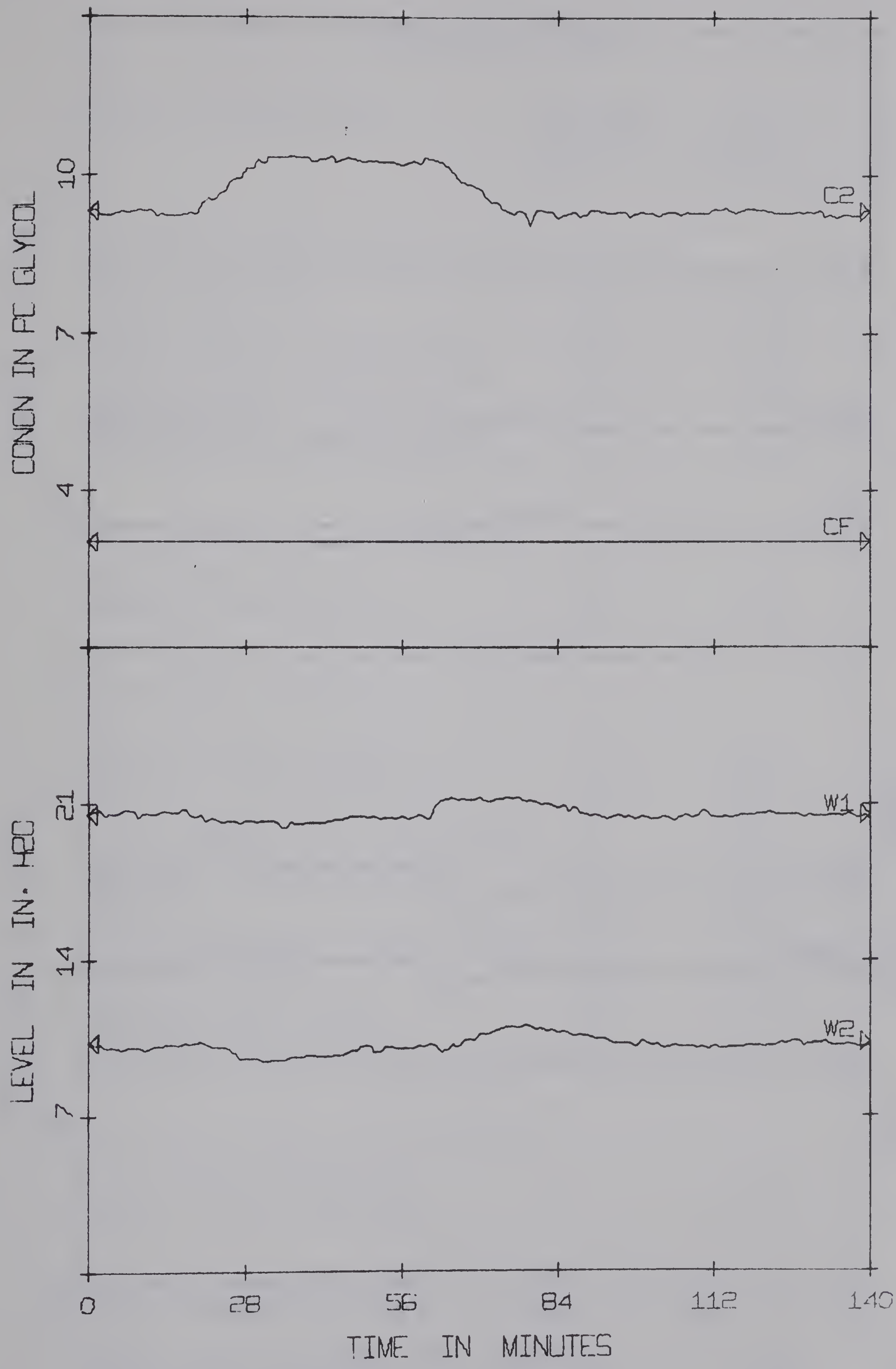
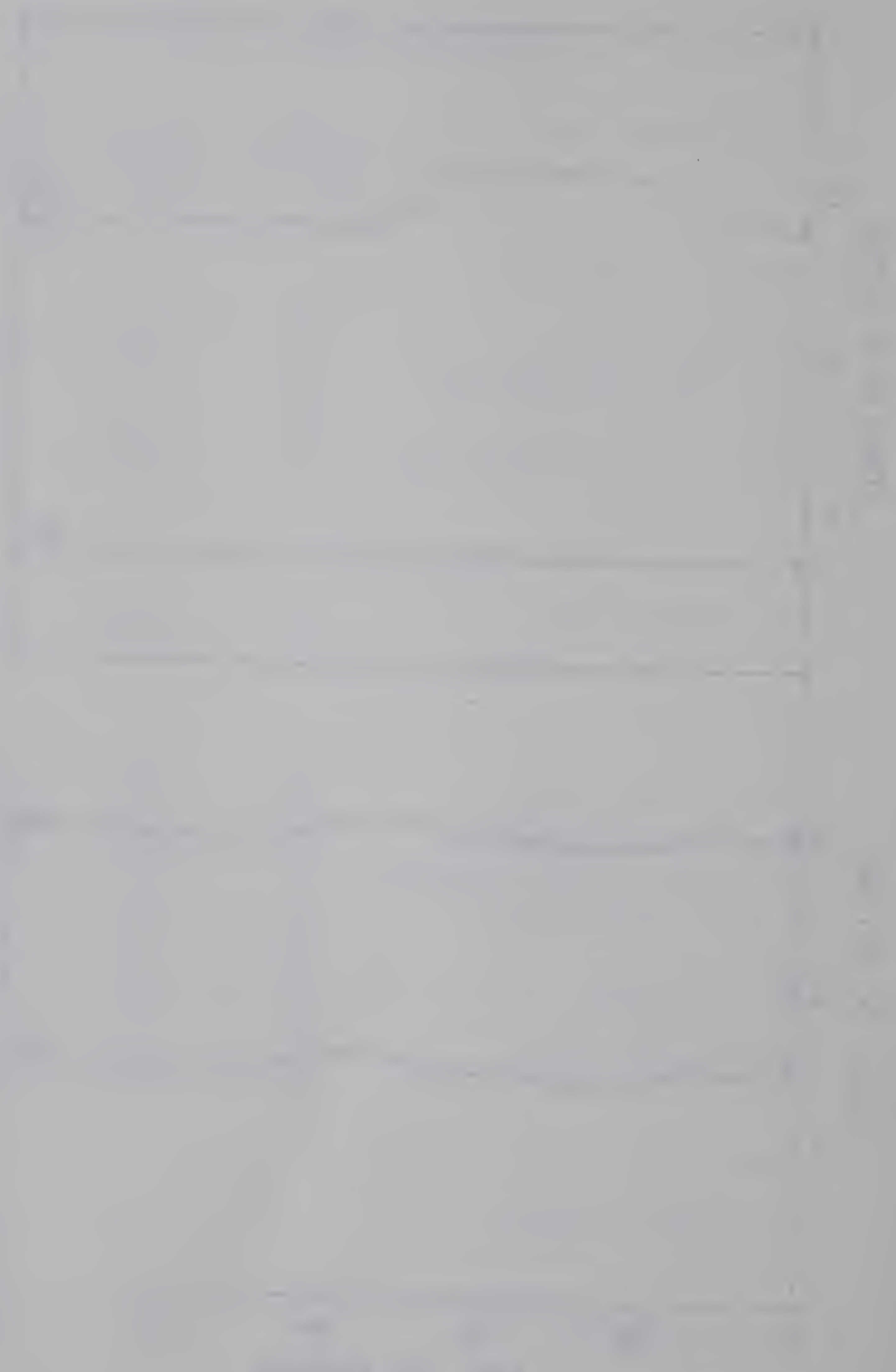


Figure A-38a Transient Data for Run LSU1 (Change C2 setpoint)



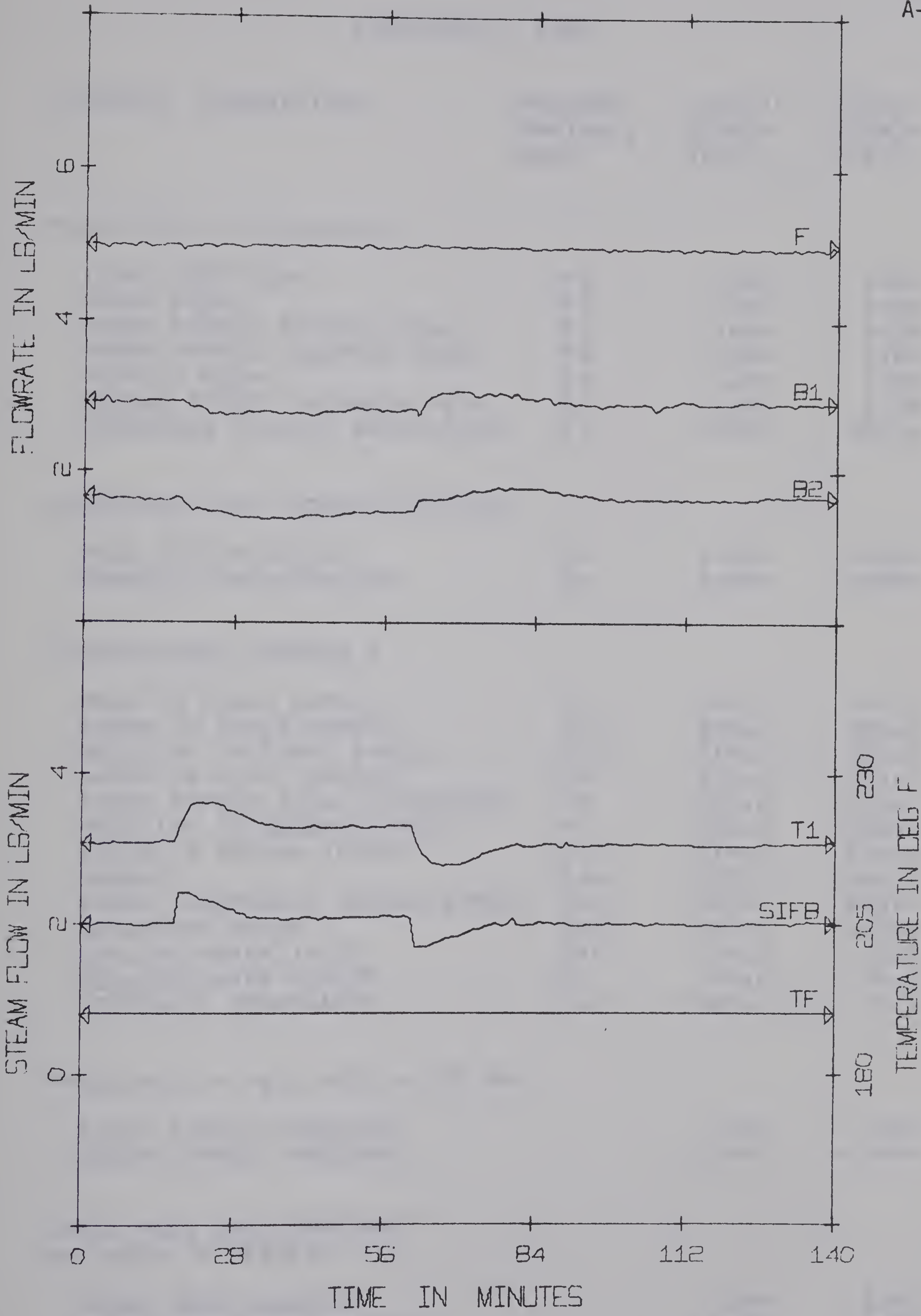


Figure A-38b Transient Data for Run LSU1 (Change C2 setpoint)

EXPERIMENT LSU2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.52	5.51
STEAM FLOW	F1	1.80	2.30
FIRST EFFECT BOTTOMS FLOW	F2	3.44	4.20
FIRST EFFECT OVERHEAD FLOW	F5	1.50	1.80
PRODUCT FLOW	F6	1.36	1.66
SECOND EFFECT OVERHEAD FLOW	F7	1.26	1.70
CONDENSER COOLING WATER FLOW	F9	40.16	40.16
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.093	0.093
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	188.7	190.3
STEAM TO FIRST EFFECT	T15	299.8	298.1
SOLUTION IN FIRST EFFECT	T19	216.1	227.6
VAPOR IN FIRST EFFECT	T2	213.9	225.5
FIRST EFFECT STEAM CONDENSATE	T5	234.0	251.4
SOLUTION TO SECOND EFFECT	T4	182.2	183.6
STEAM TO SECOND EFFECT	T10	214.4	226.0
PRODUCT	T34	157.9	160.1
STEAM CONDENSATE SECOND EFFECT	T28	193.1	208.4
SEPARATOR VAPOR	T12	161.5	162.6
COOLING WATER INLET	T29	54.8	51.9
COOLING WATER OUTLET	T1	92.6	96.8
CONDENSER CONDENSATE	T11	127.1	137.1
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		1.01	7.98
SECOND EFFECT PRESSURE		-15.48	-15.48
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		7.58	2.65
TOTAL COMPONENT BALANCE		6.90	4.16

Table A-40 Steady State Data for Run LSU2 (+20% step in F)

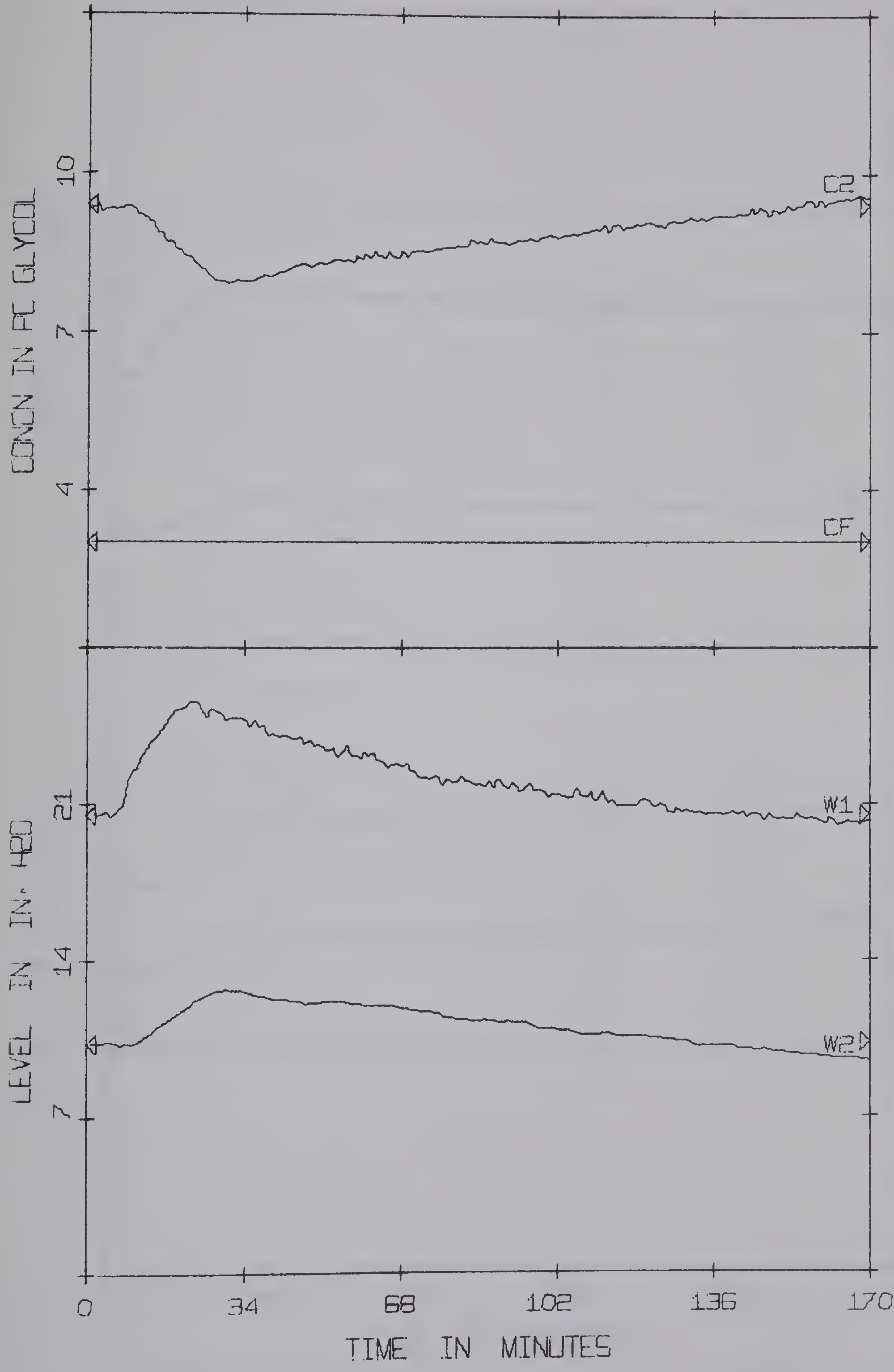


Figure A-39a Transient Data for Run LSU2 (+20% step in F)

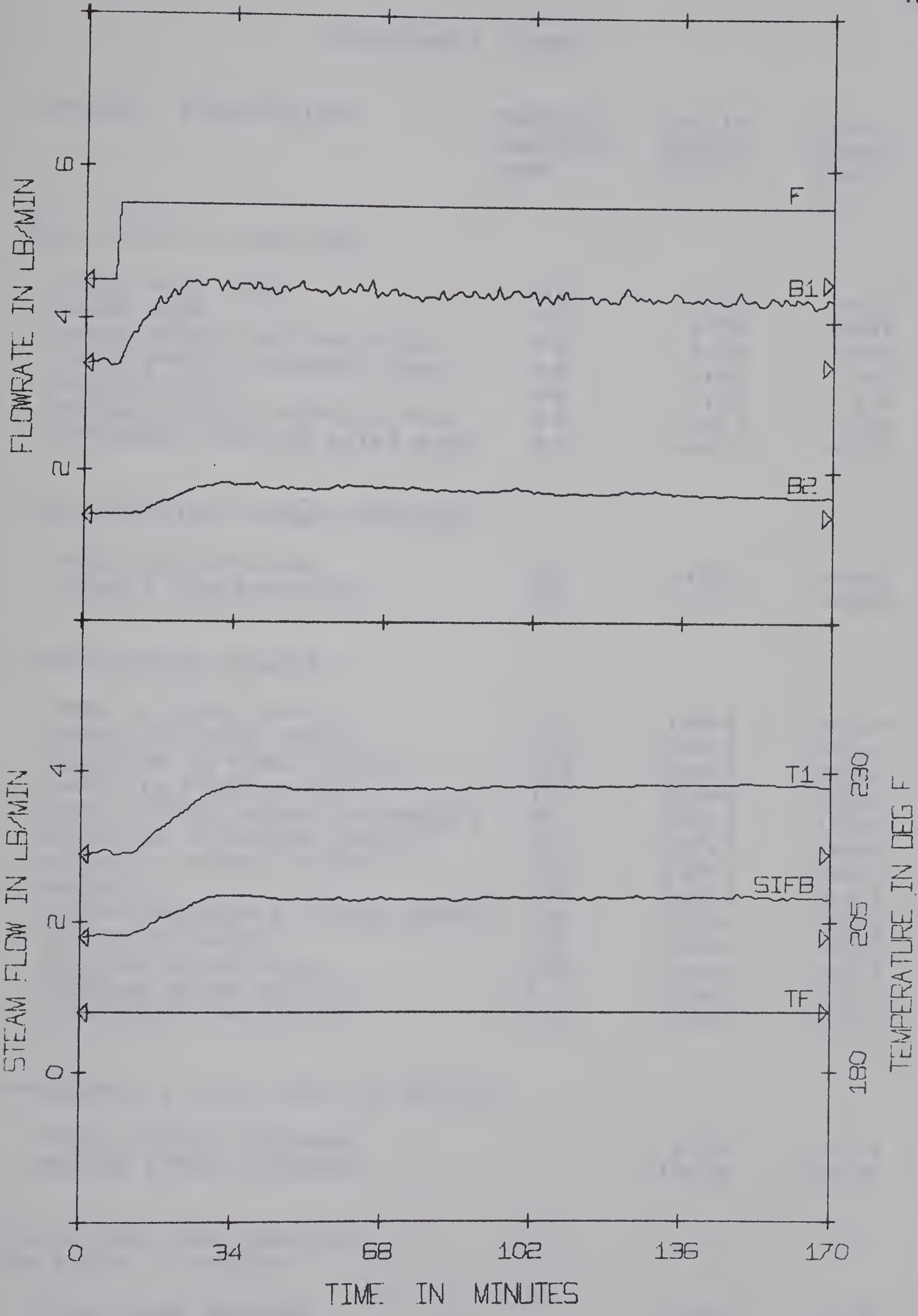


Figure A-39b Transient Data for Run LSU2 (+20% step in F)

EXPERIMENT COMB1

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.51	5.53
STEAM FLOW	F1	1.80	2.30
FIRST EFFECT BOTTOMS FLOW	F2	2.58	3.28
FIRST EFFECT OVERHEAD FLOW	F5	1.45	1.82
PRODUCT FLOW	F6	1.50	1.79
SECOND EFFECT OVERHEAD FLOW	F7	1.27	1.69
CONDENSER COOLING WATER FLOW	F9	40.00	40.00
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	189.8	191.4
STEAM TO FIRST EFFECT	T15	302.5	300.8
SOLUTION IN FIRST EFFECT	T19	221.5	232.6
VAPOR IN FIRST EFFECT	T2	219.6	230.6
FIRST EFFECT STEAM CONDENSATE	T5	242.2	257.9
SOLUTION TO SECOND EFFECT	T4	185.7	186.9
STEAM TO SECOND EFFECT	T10	219.9	231.1
PRODUCT	T34	162.3	163.6
STEAM CONDENSATE SECOND EFFECT	T28	193.4	202.0
SEPARATOR VAPOR	T12	165.3	165.3
COOLING WATER INLET	T29	66.6	68.5
COOLING WATER OUTLET	T1	103.7	114.1
CONDENSER CONDENSATE	T11	113.8	161.3
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		2.88	9.78
SECOND EFFECT PRESSURE		-15.34	-15.34
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		4.64	2.39
TOTAL COMPONENT BALANCE		0.83	3.29

Table A-41 Steady State Data for Run COMB1 (+20% step in F)

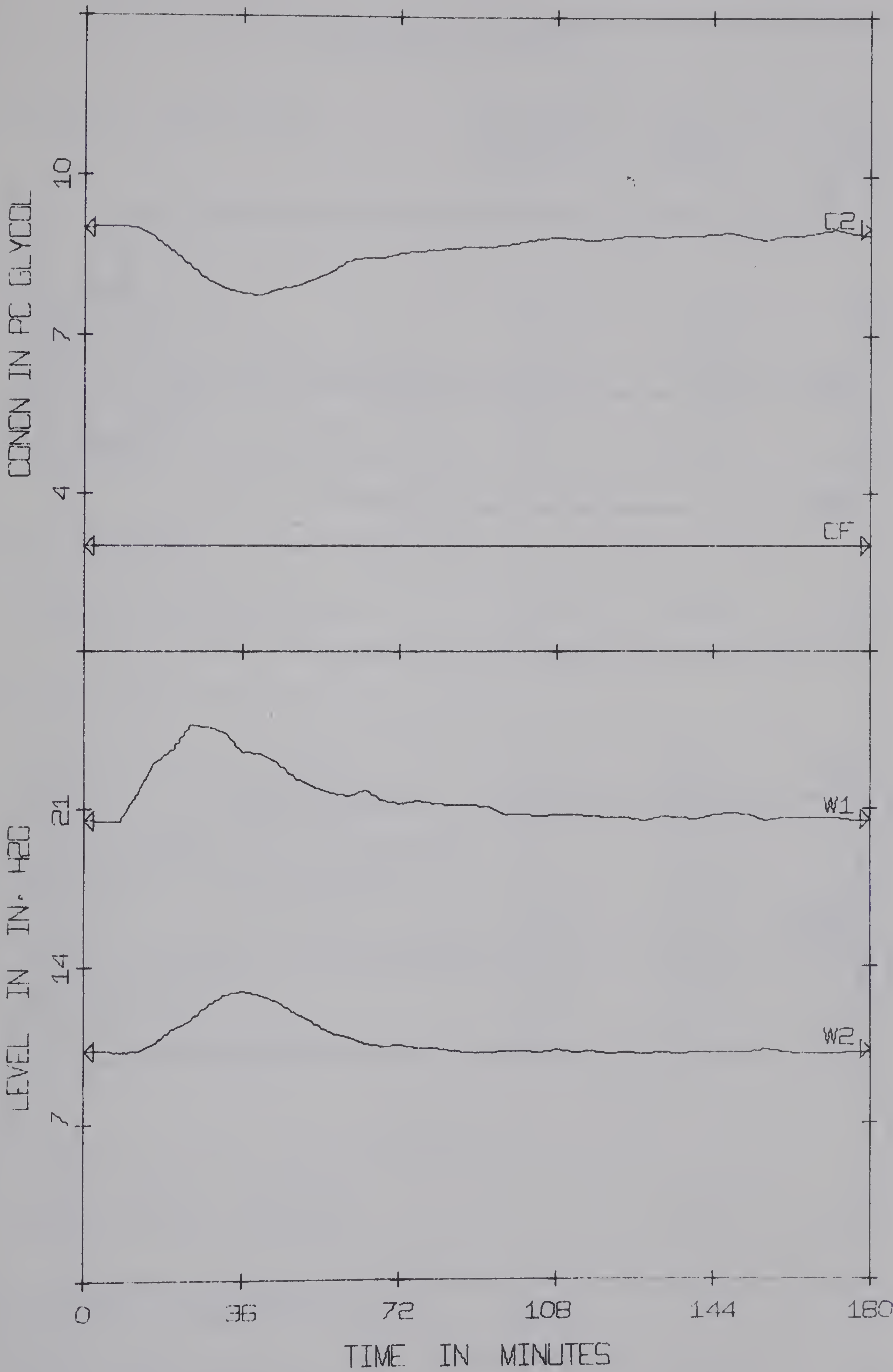


Figure A-40a Transient Data for Run COMB1 (+20% step in F)

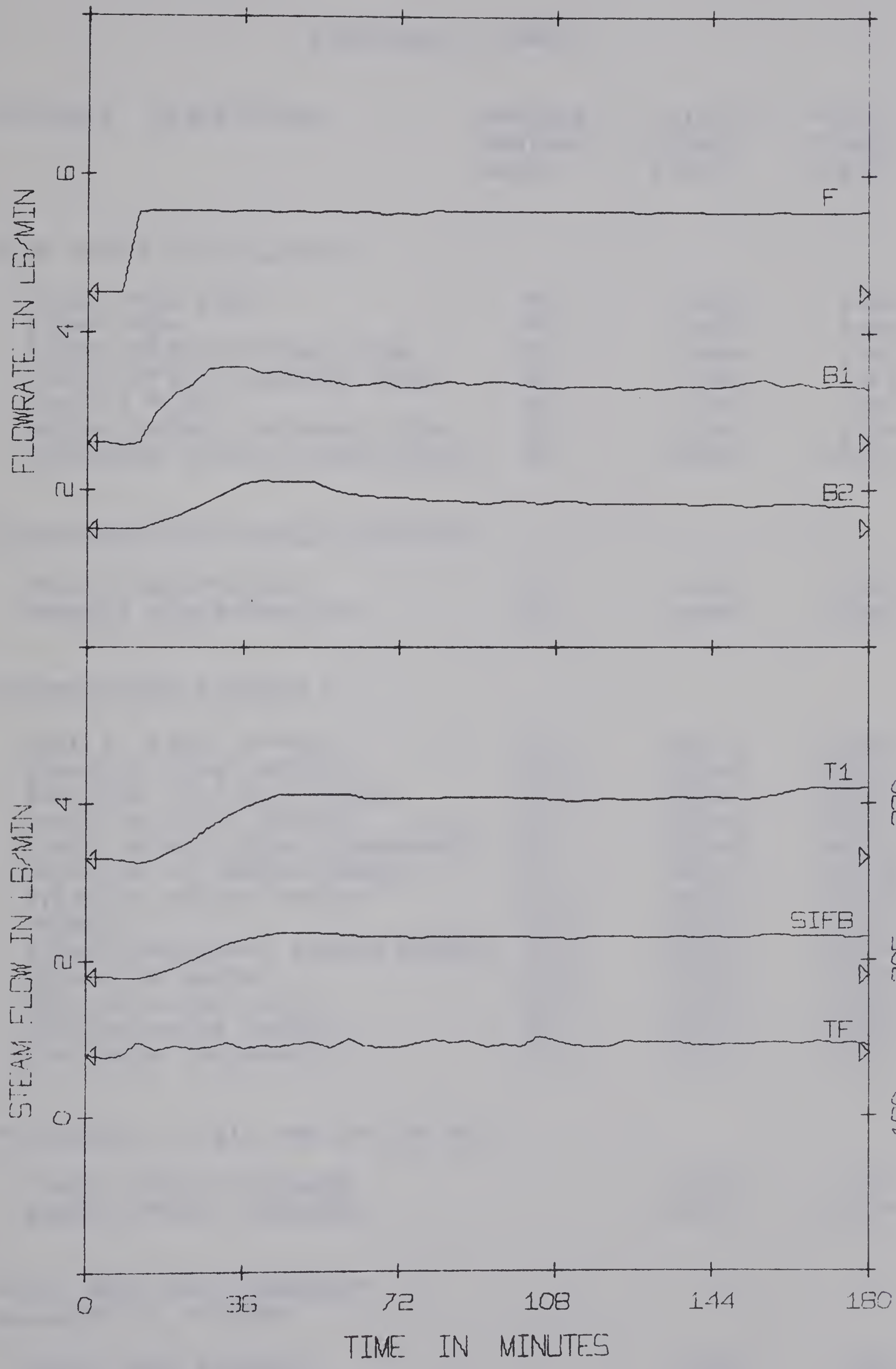


Figure A-40b Transient Data for Run COMB1 (+20% step in F)

EXPERIMENT COMB2

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	5.53	4.48
STEAM FLOW	F1	2.30	1.83
FIRST EFFECT BOTTOMS FLOW	F2	3.28	2.61
FIRST EFFECT OVERHEAD FLOW	F5	1.82	1.43
PRODUCT FLOW	F6	1.79	1.50
SECOND EFFECT OVERHEAD FLOW	F7	1.69	1.35
CONDENSER COOLING WATER FLOW	F9	40.00	40.00
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	191.4	190.1
STEAM TO FIRST EFFECT	T15	300.8	302.7
SOLUTION IN FIRST EFFECT	T19	232.6	221.8
VAPOR IN FIRST EFFECT	T2	230.6	220.1
FIRST EFFECT STEAM CONDENSATE	T5	257.9	243.0
SOLUTION TO SECOND EFFECT	T4	186.9	185.6
STEAM TO SECOND EFFECT	T10	231.1	220.5
PRODUCT	T34	163.6	162.4
STEAM CONDENSATE SECOND EFFECT	T28	202.0	193.5
SEPARATOR VAPOR	T12	165.3	165.3
COOLING WATER INLET	T29	68.5	68.8
COOLING WATER OUTLET	T1	114.1	104.9
CONDENSER CONDENSATE	T11	161.3	132.4
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		9.78	3.25
SECOND EFFECT PRESSURE		-15.34	-15.34
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		2.39	3.21
TOTAL COMPONENT BALANCE		3.29	0.34

Table A-42 Steady State Data for Run COMB2 (-20% step in F)

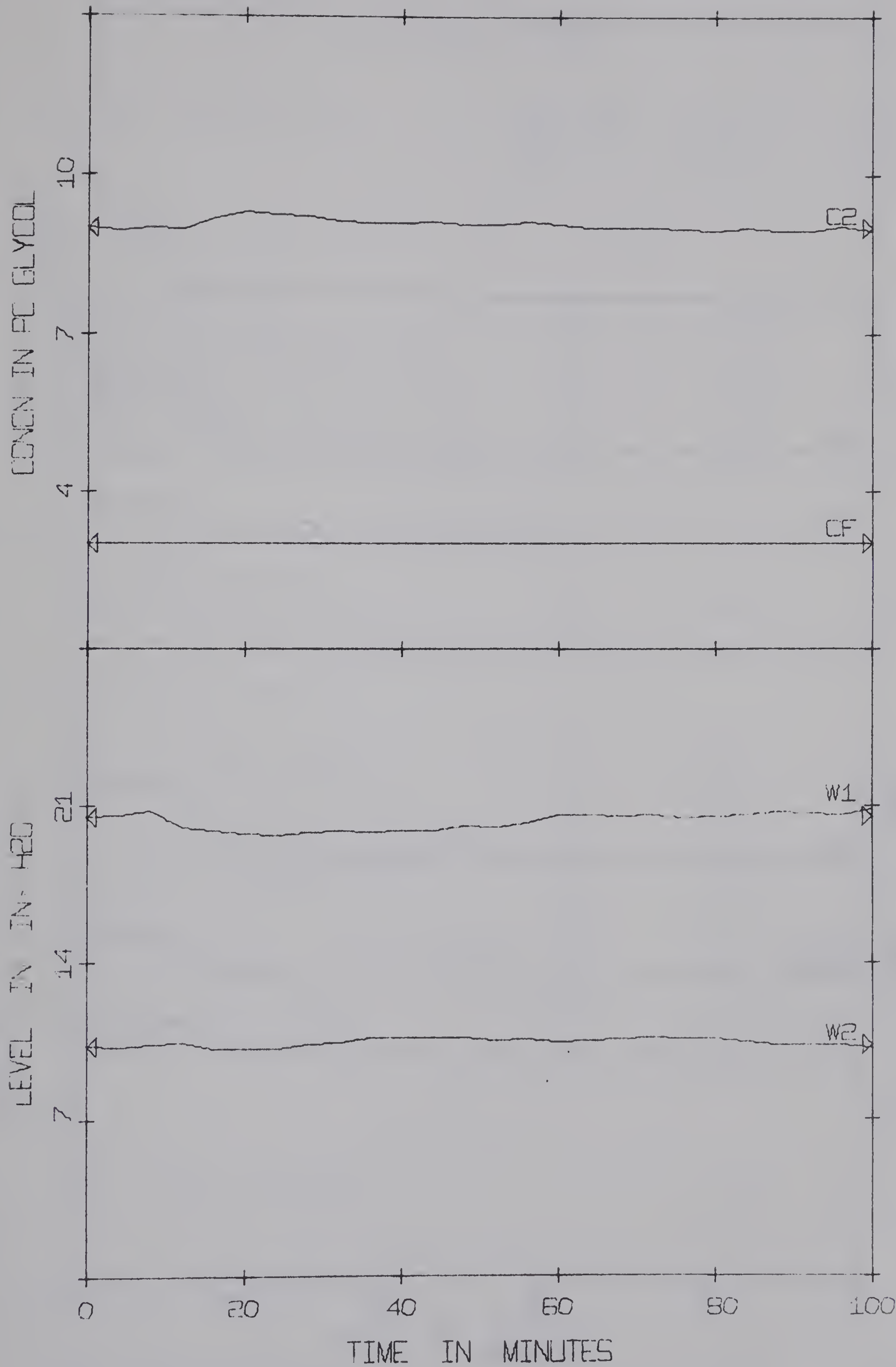


Figure A-41a Transient Data for Run COMB2 (-20% step in F)

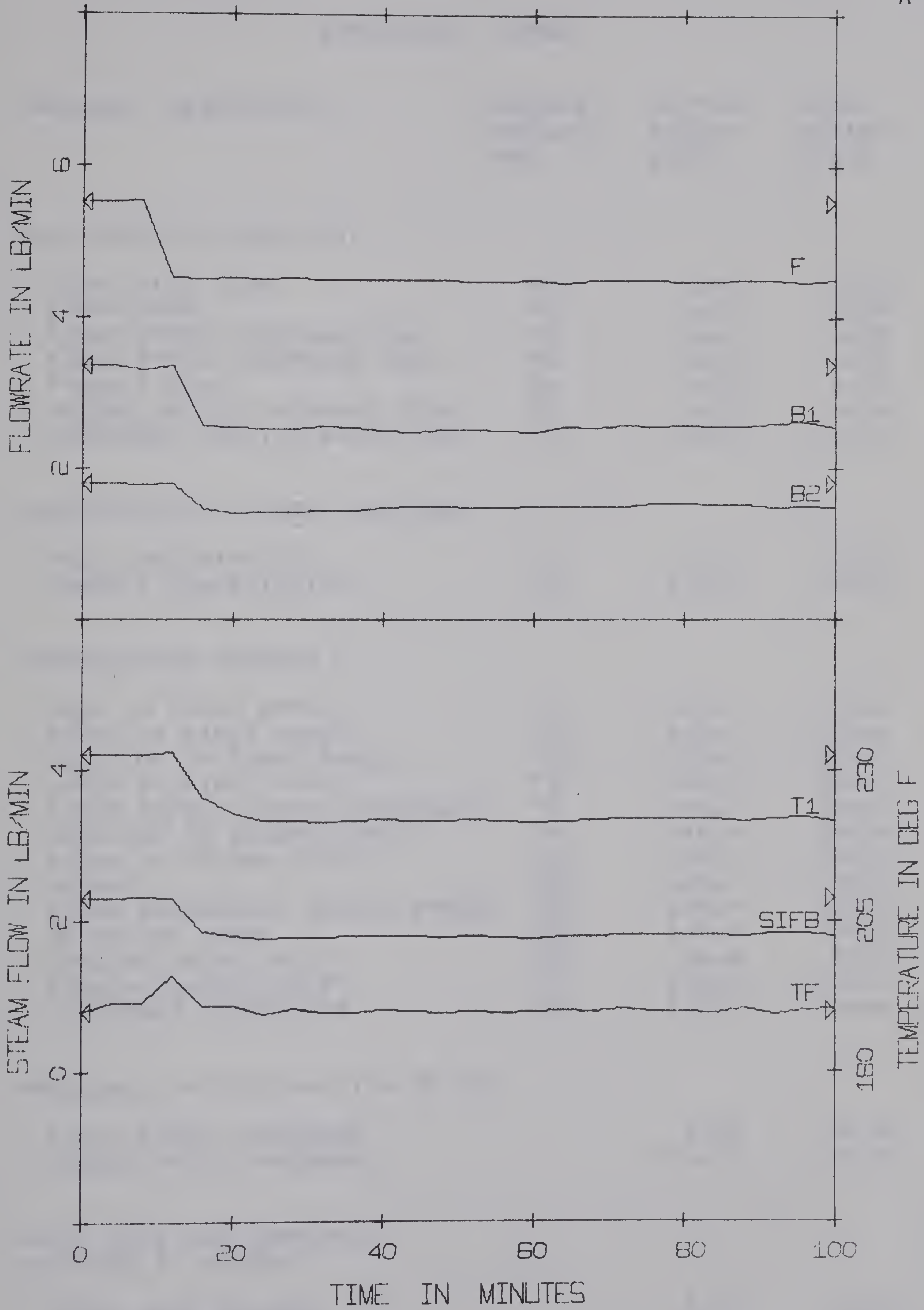


Figure A-41b Transient Data for Run COMB2 (-20% step in F)

EXPERIMENT COMB3

VARIABLE DESCRIPTION	PROCESS VARIABLE NAME	INITIAL STEADY STATE	FINAL STEADY STATE
FLOW RATES IN LBS./MIN.			
TOTAL FEED FLOW	F8	4.48	5.52
STEAM FLOW	F1	1.83	2.33
FIRST EFFECT BOTTOMS FLOW	F2	2.61	3.33
FIRST EFFECT OVERHEAD FLOW	F5	1.43	1.75
PRODUCT FLOW	F6	1.50	1.83
SECOND EFFECT OVERHEAD FLOW	F7	1.35	1.74
CONDENSER COOLING WATER FLOW	F9	40.00	40.00
CONCENTRATIONS WEIGHT FRACTION			
FEED CONCENTRATION	C1	0.030	0.030
PRODUCT CONCENTRATION	C6	0.090	0.090
TEMPERATURES DEGREES F			
FEED TO FIRST EFFECT	T7	190.1	192.2
STEAM TO FIRST EFFECT	T15	302.7	300.6
SOLUTION IN FIRST EFFECT	T19	221.8	231.9
VAPOR IN FIRST EFFECT	T2	220.1	229.9
FIRST EFFECT STEAM CONDENSATE	T5	243.0	256.7
SOLUTION TO SECOND EFFECT	T4	185.6	186.8
STEAM TO SECOND EFFECT	T10	220.5	230.3
PRODUCT	T34	162.4	163.2
STEAM CONDENSATE SECOND EFFECT	T28	193.5	202.1
SEPARATOR VAPOR	T12	165.3	165.2
COOLING WATER INLET	T29	68.8	67.7
COOLING WATER OUTLET	T1	104.9	111.7
CONDENSER CONDENSATE	T11	132.4	159.6
PRESSURES IN PSIG AND IN. OF HG.			
FIRST EFFECT PRESSURE		3.25	9.76
SECOND EFFECT PRESSURE		-15.34	-15.34
TOTAL MASS AND COMPONENT BALANCES IN PERCENT			
TOTAL MASS BALANCE		3.21	1.92
TOTAL COMPONENT BALANCE		0.34	1.66

Table A-43 Steady State Data for Run COMB3 (+20% step in F)

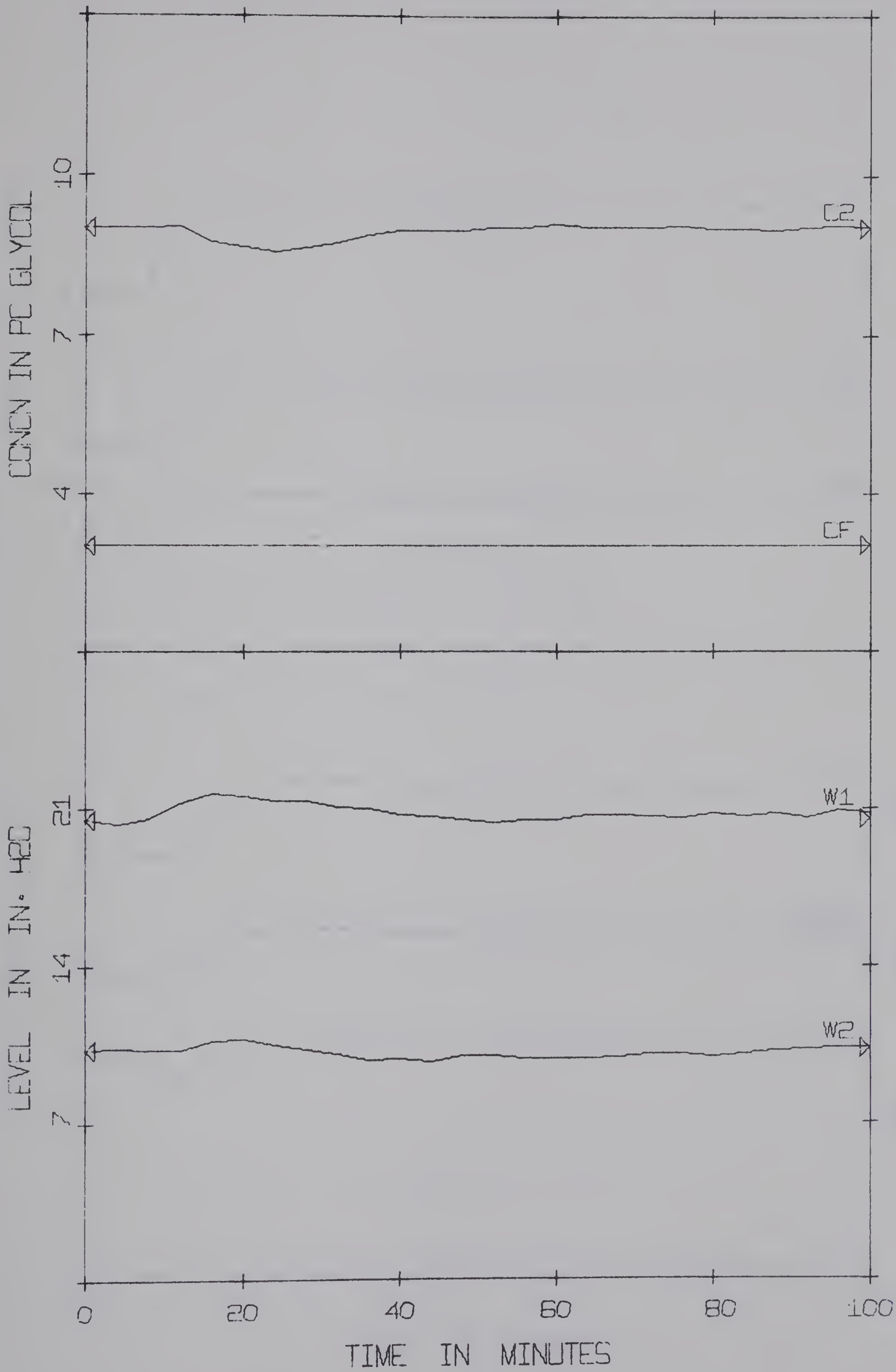


Figure A-42a Transient Data for Run COMB3 (+20% step in F)

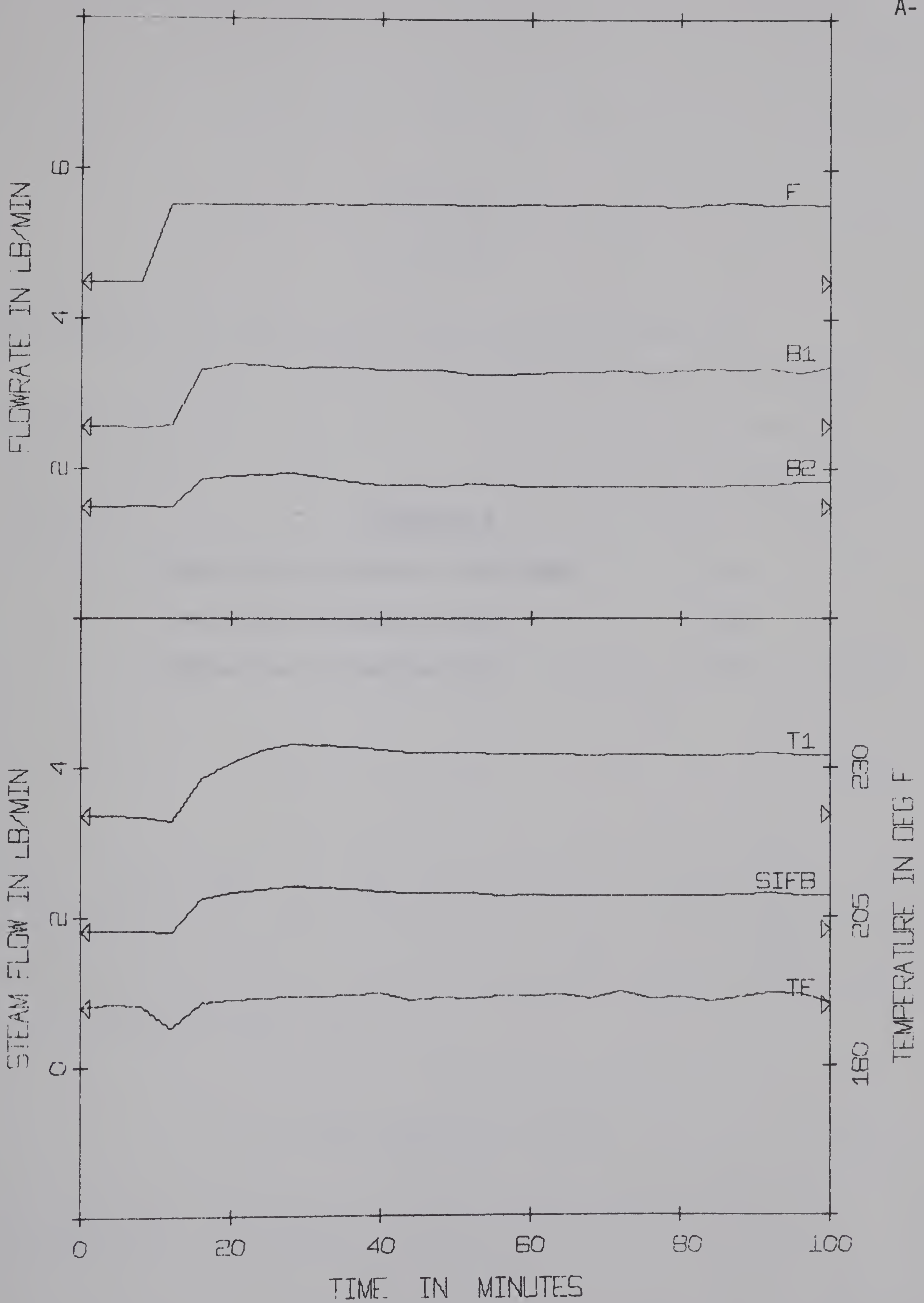


Figure A-42b Transient Data for Run COMB3 (+20% step in F)

APPENDIX B

APPENDIX B

DERIVATIONS

Derivation of the Seventh Order Linearized Model (Table 3.4)

In order to incorporate the level controllers on W1 and W2 into the mathematical model, the disturbance vector is neglected and only B1 and B2, W1 and W2 are used in the manipulated and output vectors, respectively. Therefore the fifth order model is modified and has the following form:

$$\dot{\underline{X}} = \underline{\underline{A}}\,\underline{X} + \underline{\underline{B1}}\,\underline{U} \tag{B-1a}$$

$$\underline{Y} = \underline{\underline{E1}}\,\underline{X} \tag{B-1b}$$

where $\underline{\underline{B1}}$ and $\underline{\underline{E1}}$ are the modified $\underline{\underline{B}}$ and $\underline{\underline{E}}$ coefficient matrices. Assuming the initial steady states used in the model are equivalent to the desired setpoints for W1 and W2, then the manipulated variables are given by the PID control law:

$$\underline{U} = \underline{\underline{KP}}\,\underline{Y} + \underline{\underline{KI}} \int \underline{Y}\,dt + \underline{\underline{KD}}\,\dot{\underline{Y}} \tag{B-2}$$

since the model variables are in normalized perturbation form. Let

$$\underline{Z} = \int \underline{Y}\,dt \tag{B-3}$$

Therefore

$$\dot{\underline{Z}} = \underline{Y} = \underline{\underline{E1}}\ \underline{X} \tag{B-4}$$

And

$$\underline{U} = \underline{\underline{KP}}\ \underline{\underline{E1}}\ \underline{X} + \underline{\underline{KI}}\ \underline{Z} + \underline{\underline{KD}}\ \underline{\underline{E1}}\ \dot{\underline{X}} \tag{B-5}$$

Combine equation B-1a and B-5 and solve for $\dot{\underline{X}}$

$$\dot{\underline{X}} = (\underline{\underline{I}} - \underline{\underline{B1}}\ \underline{\underline{KD}}\ \underline{\underline{E1}})^{-1}\ [(\underline{\underline{A}} + \underline{\underline{B1}}\ \underline{\underline{KP}}\ \underline{\underline{E1}})\underline{X} + \underline{\underline{B1}}\ \underline{\underline{KI}}\ \underline{Z}] \tag{B-6}$$

The seventh order model can be written using equations B-4 and B-6, and including the load (disturbance) vector, $(F\ CF\ hF)^T$ and the remaining manipulated variable, SIFB:

$$\begin{bmatrix} \dot{\underline{X}} \\ \dot{\underline{Z}} \end{bmatrix} = \begin{bmatrix} (\underline{\underline{I}} - \underline{\underline{B1}}\ \underline{\underline{KD}}\ \underline{\underline{E1}})^{-1}(\underline{\underline{A}} + \underline{\underline{B1}}\ \underline{\underline{KP}}\ \underline{\underline{E1}}) & (\underline{\underline{I}} - \underline{\underline{B1}}\ \underline{\underline{KD}}\ \underline{\underline{E1}})^{-1}\underline{\underline{B1}}\ \underline{\underline{K1}} \\ \underline{\underline{E1}} & \underline{\underline{0}} \end{bmatrix} \begin{bmatrix} \underline{X} \\ \underline{Z} \end{bmatrix} + \tag{B-7a}$$

$$\underline{\underline{B2}}\ \underline{U} + \underline{\underline{C1}}\ \underline{D}$$

$$\underline{Y} = \underline{\underline{E2}} \begin{bmatrix} \underline{X} \\ \underline{Z} \end{bmatrix} \tag{B-7b}$$

or defining new state variables by $[\underline{X}\ \underline{Z}]^T$

$$\dot{\underline{X}} = \underline{\underline{A1}}\ \underline{X} + \underline{\underline{B2}}\ \underline{U} + \underline{\underline{C1}}\ \underline{D} \tag{B-8a}$$

$$\underline{Y} = \underline{\underline{E2}}\ \underline{X} \tag{B-8b}$$

where $\underline{X} = [W1\ C1\ h1\ W2\ C2\ X6\ X7]^T$

$\underline{U} = [SIFB]$

$\underline{D} = [F\ CF\ hF]^T$

$\underline{Y} = [W1\ W2\ C2]^T$

Derivation of the Coefficient Matrix in Equation 5.2b

After a setpoint change in the output variables with proportional control only, the model difference equations, in normalized perturbation form, are given by the following at the new steady state.

$$\underline{XSS} = \underline{\underline{\Phi}}\ \underline{XSS} + \underline{\underline{H1}}\ \underline{USS} \tag{B-9a}$$

$$\underline{YSS} = \underline{\underline{E}}\ \underline{XSS} \tag{B-9b}$$

where

$$\underline{USS} = \underline{\underline{KP}}\ (\underline{YSS} - \underline{R}) \tag{B-10}$$

or

$$\underline{USS} = \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{XSS} - \underline{\underline{KP}}\ \underline{R} \tag{B-11}$$

Substitute equation B-11 into B-9a and solve for $\underline{XSS}$

$$\underline{XSS} = - (\underline{\underline{I}} - \underline{\underline{\Phi}} - \underline{\underline{H1}}\ \underline{\underline{KP}}\ \underline{\underline{E}})^{-1}\ \underline{\underline{H1}}\ \underline{\underline{KP}}\ \underline{R} \qquad \text{(B-12a)}$$

or

$$\underline{YSS} = - E(\underline{\underline{I}} - \underline{\underline{\Phi}} - \underline{\underline{H1}}\ \underline{\underline{KP}}\ \underline{\underline{E}})^{-1}\underline{\underline{H1}}\ \underline{\underline{KP}}\ \underline{R} \qquad \text{(B-12b)}$$

Solve

$$\underline{R} = \underline{\underline{K}}\ \underline{YSS} = \underline{\underline{K}}\ \underline{SP} \qquad \text{(B-13)}$$

where

$$\underline{\underline{K}} = - [\underline{\underline{E}}(\underline{\underline{I}} - \underline{\underline{\Phi}} - \underline{\underline{H1}}\ \underline{\underline{KP}}\ \underline{\underline{E}})^{-1}\ \underline{\underline{H1}}\ \underline{\underline{KP}}]^{-1} \qquad \text{(B-14)}$$

<u>Derivation of the Combined Control Algorithm (Equation 5.5)</u>

The manipulated variables at time interval n are given by

$$\underline{U}(n) = \underline{UP}(n) + \underline{UI}(n) + \underline{UFF}(n) \qquad \text{(B-15)}$$

where $\underline{UP}$, $\underline{UI}$ and $\underline{UFF}$ are the proportional, integral and feed forward contributions, respectively.

Each contribution is derived below in normalized perturbation units.

(a) Proportional Control - $\underline{UP}$

The proportional control law at time n is

$$\underline{UP}(n) = \underline{\underline{KP}}\ (\underline{YC}(n+\tfrac{1}{2}) - \underline{R}(n)) = \underline{\underline{KP}}\ \underline{YC}(n+\tfrac{1}{2}) - \underline{\underline{KP}}\ \underline{\underline{K}}\ \underline{SP}(n) \qquad \text{(B-16)}$$

with one-half sampling time prediction on the output variables.

For linear prediction into the future

$$\underline{YC}(n+\tfrac{1}{2}) = .5\ \underline{YS}(n) + .5\ \underline{YC}(n+1) \qquad \text{(B-17)}$$

where $$\underline{YS}(n) = \underline{YC}(n) + \underline{\underline{AY}}\ (\underline{YM}(n) - \underline{YC}(n)) \qquad \text{(B-18)}$$

and $$\underline{YC}(n+1) = \underline{\underline{E}}\ \underline{XC}(n+1) + \underline{\underline{F}}\ \underline{VM}(n) \qquad \text{(B-19)}$$

and $$\underline{XC}(n+1) = \underline{\underline{\Phi}}\ \underline{XS}(n) + \underline{\underline{H1}}\ \underline{U}(n) + \underline{\underline{H2}}\ \underline{DM}(n) \qquad \text{(B-20)}$$

Since both YS(n) and XS(n) are calculated in the model section at time n, they are available for use in the control section and therefore these variables will be retained throughout the derivation. Combining equations B-16, B-17, B-19 and B-20 results in

$$\underline{UP}(n) = .5\ \underline{\underline{KP}}\ \underline{YS}(n) + .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{\Phi}}\ \underline{XS}(n) + .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}}\ \underline{U}(n) + .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H2}}\ \underline{DM}(n)$$

$$+ .5\ \underline{\underline{KP}}\ \underline{\underline{F}}\ \underline{VM}(n) - \underline{\underline{KP}}\ \underline{\underline{K}}\ \underline{SP}(n) \qquad \text{(B-21)}$$

(b) Integral Control - $\underline{UI}$

The integral control law is given by

$$\underline{UI}(n) = \underline{UI}(n-1) + \underline{\underline{KI}}\ \underline{E}(n) \qquad \text{(B-22)}$$

where $\underline{E}(n)$ represents either normal or the modified integral mode by using

$$\underline{E}(n) = \underline{YS}(n) - \underline{\hat{YC}}(n) - \underline{\underline{AI}}\ (\underline{SP}(n) - \underline{\hat{YC}}(n)) \qquad \text{(B-23)}$$

$$\hat{Y}C(n) = \underline{XC}(n) - \underline{\underline{H1}}\ (\underline{UI}(n-1) \qquad \text{(B-24)}$$

Therefore

$$\underline{UI}(n) = (\underline{\underline{I}} + \underline{\underline{KI}}\ \underline{\underline{H1}} - \underline{\underline{KI}}\ \underline{\underline{AI}}\ \underline{\underline{H1}})\ \underline{UI}(n-1) + \underline{\underline{KI}}\ \underline{YS}(n)$$

$$+ (\underline{\underline{KI}}\ \underline{\underline{AI}} - \underline{\underline{KI}})\ \underline{XC}(n) - \underline{\underline{KI}}\ \underline{\underline{AI}}\ \underline{SP}(n) \qquad \text{(B-25)}$$

(c) Feed Forward Control - $\underline{UFF}$

The feed forward control law is given by

$$\underline{UFF}(n) = \underline{\underline{KFF}}\ \underline{DM}(n) \qquad \text{(B-26)}$$

Substitute B-21, B-25 and B-26 into B-15 and solve for $\underline{U}(n)$

$$\underline{U}(n) = \underline{\underline{K1}}\ \underline{YS}(n) + \underline{\underline{K2}}\ \underline{XS}(n) + \underline{\underline{K3}}\ XC(n) + \underline{\underline{K4}}\ \underline{DM}(n) + \underline{\underline{K5}}\ \underline{VM}(n)$$

$$+ \underline{\underline{K6}}\ \underline{SP}(n) + \underline{\underline{K7}}\ \underline{UI}(n-1) \qquad \text{(B-27)}$$

where

$$\underline{\underline{K1}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (.5\ \underline{\underline{KP}} + \underline{\underline{KI}})$$

$$\underline{\underline{K2}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (.5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{\Phi}})$$

$$\underline{\underline{K3}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (\underline{\underline{KI}}\ \underline{\underline{AI}} - \underline{\underline{KI}})$$

$$\underline{\underline{K4}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (.5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H2}} + \underline{\underline{KFF}})$$

$$\underline{\underline{K5}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (.5\ \underline{\underline{KP}}\ \underline{\underline{F}})$$

$$\underline{\underline{K6}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (-\ \underline{\underline{KP}}\ \underline{\underline{K}} - \underline{\underline{KI}}\ \underline{\underline{AI}})$$

$$\underline{\underline{K7}} = (\underline{\underline{I}} - .5\ \underline{\underline{KP}}\ \underline{\underline{E}}\ \underline{\underline{H1}})^{-1}\ (\underline{\underline{I}} + \underline{\underline{KI}}\ \underline{\underline{H1}} - \underline{\underline{KI}}\ \underline{\underline{AI}}\ \underline{\underline{H1}})$$

APPENDIX C

Table C-1

List of DDC Loop Records

Loop	Name	COS	MPX	Type	Description
0125	L15		28	C	Second effect condensate level
0126	F7		42	I	Separator overhead flow
0129	C6		35	I	Product concentration
0131	F2			C	Feed forward controller
0132	F1	III-6	0	C	Steam to first effect
0133	F9	II-6	16	C	Condenser cooling water
0134	P22	III-3	25	C	Condenser vacuum
0135	P20		41	I	First effect pressure
0136	L14		22	C	First effect liquid level
0137	L11		19	C	Separator liquid level
0138	TT10	II-5	84	C	Feed solution temperature
0139	F12	II-4	9	C	Feed solution flow
0141	F11	II-3	6	C	Feed water flow
0142	TT11	II-2	83	C	Feed water temperature
0143	C1		32	I	Feed concentration
0144	C6		35	C	Product concentration
0145	P20	III-4	41	C	First effect pressure
0149	F8		12	I	Total feed flow
0150	F5	III-5	38	C	Second effect condensate flow
0151	F2	III-2	39	C	First effect bottoms flow
0152	F6	III-1	40	C	Product flow
0153			65	I	Bridge unbalance measurement
0154			64	C	Reference voltage
0159	T7		91	I	Feed to first effect
0163	T34		79	I	Product from second effect
0164	T19		89	I	Liquid in first effect

COS - current output station
MPX - relay multiplexer point
C - control loop
I - data acquisition loop

Table C-2

Listing of the Process Variable Table

DDC Control Loops

DDC Loop 0125

```
012501   621E  104E+00028  3400  9588+01715+00158+03644+24533+32767
012502+32767   3000+32767+00000  1110+32767+00000  1550  0150+00000
012503   40A1  0000+09783+21824  0004  0200  0000  0000+00000
```

DDC Loop 0131

```
013101   821E  607E  0151  0000  8488+12480  0000  0000  0000+32767
013102+32767   1000+32767  0000  1110+32767-32767  1550  0144  0000
013103   00A2  0000  0000  0000  0000  0075  0000  0000  0000
```

DDC Loop 0132

```
013201   421E  105E+00000  0000  8488+08349+00000+00000+04940+32767
013202+32767   3000+32767+00000  1110+32767+00000  1550  1D78+00982
013203   03C0  0000+00000+00000  0100  0020  0000  0000+00000
```

DDC Loop 0133

```
013301   621E  005F+00016  3400  9488+10080+00000+00000+27306+32767
013302+32767   3000+32767+00000  1110+32767+00000  1550  1578+27845
013303   40C0  0000+32767+21312  0001  0800  0000  0000+00000
```

DDC Loop 0134

```
013401   421E  104F+00025  0000  9688+06800-03000+12356+15420+32767
013402+32767   3000+32767+00000  1110+32767+00000  1550  1A78+00000
013403   40C0  0000+21610+09152  0001  4C00  0000  0000+00000
```

DDC Loop 0136

```
013601   621E  104E+00022  3000  9588+06300+00000+13432+20805+32767
013602+32767   3000+32767+00000  1110+32767+00000  1550  0151+11733
013603   40A1  0000+12491-07744  0001  0080  0000  0000+00000
```

DDC Loop 0137

```
013701   621E  104E+00019  3000  9588+05000+00000+10005+14417+32767
013702+32767   3000+32767+00000  1110+32767+00000  1550  0152+15913
013703   40A1  0000+17866+18304  0001  0180  0000  0000+00000
```

DDC Loop 0138

```
013801   621E  10CE+00084  3000  A288+20000+00000+03578+06225+32767
013802+32767   3000+32767+00000  1110+32767+00000  1550  1478+00000
013803   40C0  0000-32768-32768  0100  0080  0000  0000+00000
```

Table C-2 (continued)

```
DDC Loop  0139

  013901  521E  005E+00009  0800  8488+06100+00000+00000+00000+32767
  013902+32767  1000+32767+00000  1110+32767+00000  1550  1378+00000
  013903  20C0  0000+00000+00000  0200  0038  0000  0000+00000

DDC Loop 0141

  014101  521E  005E+00006  0800  8488+10200+00000+00000+00000+32767
  014102+32767  1000+32767+00000  1110+32767+00000  1550  1278+00000
  014103  22C0  0000+00000+00000  0040  0060  0000  0000+00000

DDC Loop  0142

  014201  621E  10CE+00083  3000  A288+20000+00000+03292+06225+32767
  014202+32767  3000+32767+00000  1110+32767+00000  1550  1178+05194
  014203  20C0  0000+05210+00000  0200  0080  0000  0000+00000

DDC Loop  0144

  014401  021E  600E+00000  0000  C088+03820+00000+13750+13750+32767
  014402+32767  3000+32767+00000  1110+32767+00000  1550  0132+04913
  014403  00E1  0000+17468+00000  0018  0190  0000  0000+00000

DDC Loop  0145

  014501  621E  104E+00041  3000  9688+10960-01830+00000+13932+32767
  014502+32767  3000+32767+00000  1110+32767+00000  1550  1B78+22931
  014503  20C0  0000+00000+00000  0000  0380  0000  0000+00000

DDC Loop  0150

  015001  521E  005F+00038  0800  8488+06460+00000+03115+00000+32767
  015002+32767  3000+32767+00000  1110+32767+00000  1550  1C78+06551
  015003  43C0  0000+03960-29952  0040  0080  0000  0000+0000

DDC Loop  0151

  015101  521E  005F+00039  0800  8488+12480+00000+11819+11513+32767
  015102+32767  3000+32767+00000  1110+32767+00000  1550  1978+19655
  015103  4380  0000+27233+29568  0040  0060  0000  0000+00000

DDC Loop  0152

  015201  521E  005F+00040  0800  8488+05340+00000+16648+15838+32767
  015202+32767  3000+32767+00000  1110+32767+00000  1550  1878+16379
  015203  4380  0000+05666=22272  0040  0038  0000  0000+00000

DDC Loop  0154

  015401  6F16  6B0C+00064  3200  9720+04000+00000+13169+00000+32767
  015402+32767  3220+32767-32768  3220+32767-32768  3220  0153+00000
  015403  F320
```

Table C-2 (continued)

Data Acquisition Loops

DDC Loop 0126

```
012601  600D  115F+00042  3200  8408+07500+00000+07286+00000+32767
012602+32767  3000
```

DDC Loop 0129

```
012901  600D  114F  00035  3000  C008+03820+00000+00000+00000+32767
012902+32767  1000
```

DDC Loop 0135

```
013501  600D  114F+00041  3000  9608+10960=01830+00000+00000+32767
013502+32767  1000
```

DDC Loop 0143

```
014301  000D  110F+00000  0000 C008+01200+00000+00000+00000+32767
014302+32767  1000
```

DDC Loop 0149

```
014901  500D  115F+00012  7FF0  8408+17440+00000+00000+00000+32767
014902+32767  1000
```

DDC Loop 0153

```
015301  600D  6C0C+00065  3200  9700+04000+00000+08718+00000+32767
015302+32767  3220
```

DDC Loop 0159

```
015901  600D  11CF+00091  3400  A208+20000+00000+03444+00000+32767
015902+32767  1000
```

DDC Loop 0163

```
016301  600D 11CF+00079  3400  A208+20000  0000  0000  0000+32767
016302 32767 3000
```

DDC Loop 0164

```
016401  600D  11CF+00089  3400  A208+20000+00000+06849+00000+32767
016402+32767  1000
```

Table C-3

Typical DDC Control Loops Filter and Control Constants

DDC Control Loop ID	Process Variable	Filter* Type	Filter* Constant	Poll Time (sec)	Proportional Constant	Integral Constant (sec)
0125	L15	Exp	.8125	2	4.0	512.0
0132	F1	None	--	2	0.25	8.0
0133	F9	Exp	.8125	1	1.0	1024.0
0134	P22	None	--	2	152.0	2048.0
0136	L14	Exp	.750	2	1.0	2048.0
0137	L11	Exp	.750	2	3.0	2048.0
0138	TT10	Exp	.750	2	1.0	8.0
0139	F12	Union	128	1	0.44	12.0
0141	F11	Union	128	1	0.75	16.0
0142	TT11	Exp	.750	2	1.0	4.0
0144	C6	Exp	.750	64	3.125	2728.0
0145	P20	Exp	.750	2	7.0	--
0150	F5	Union	128	1	1.0	16.0
0151	F2	Union	128	1	0.75	16.0
0152	F6	Union	128	1	0.44	16.0

*See Ref. (22), (3), (23)

Exp - exponential filter

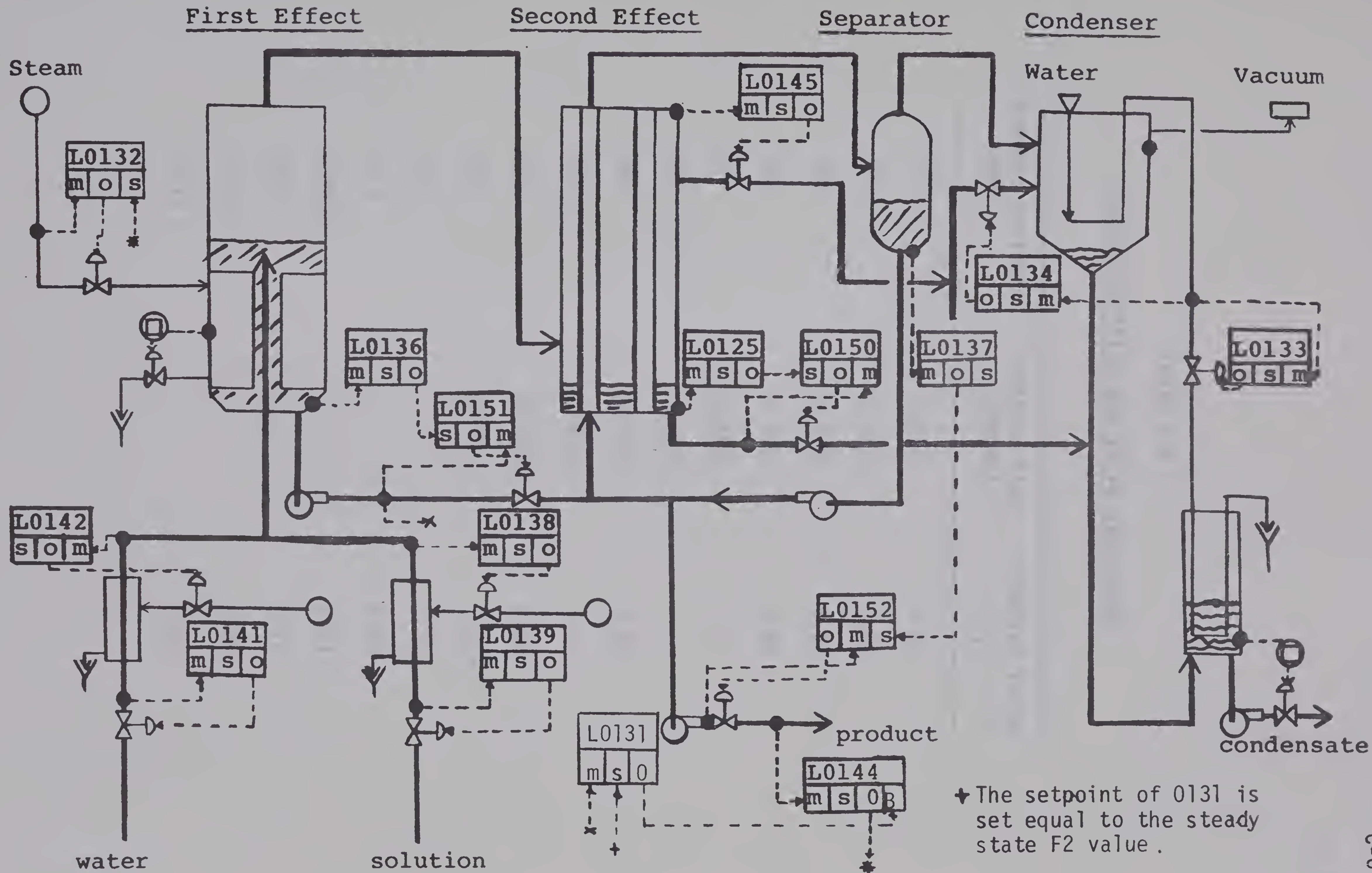


Figure C-1 DDC Feedforward and Feedback Concentration Controller for Run FF1 (Addition of Loop 0131 and the Bias on 0144 are the only changes from Figure 2.2)

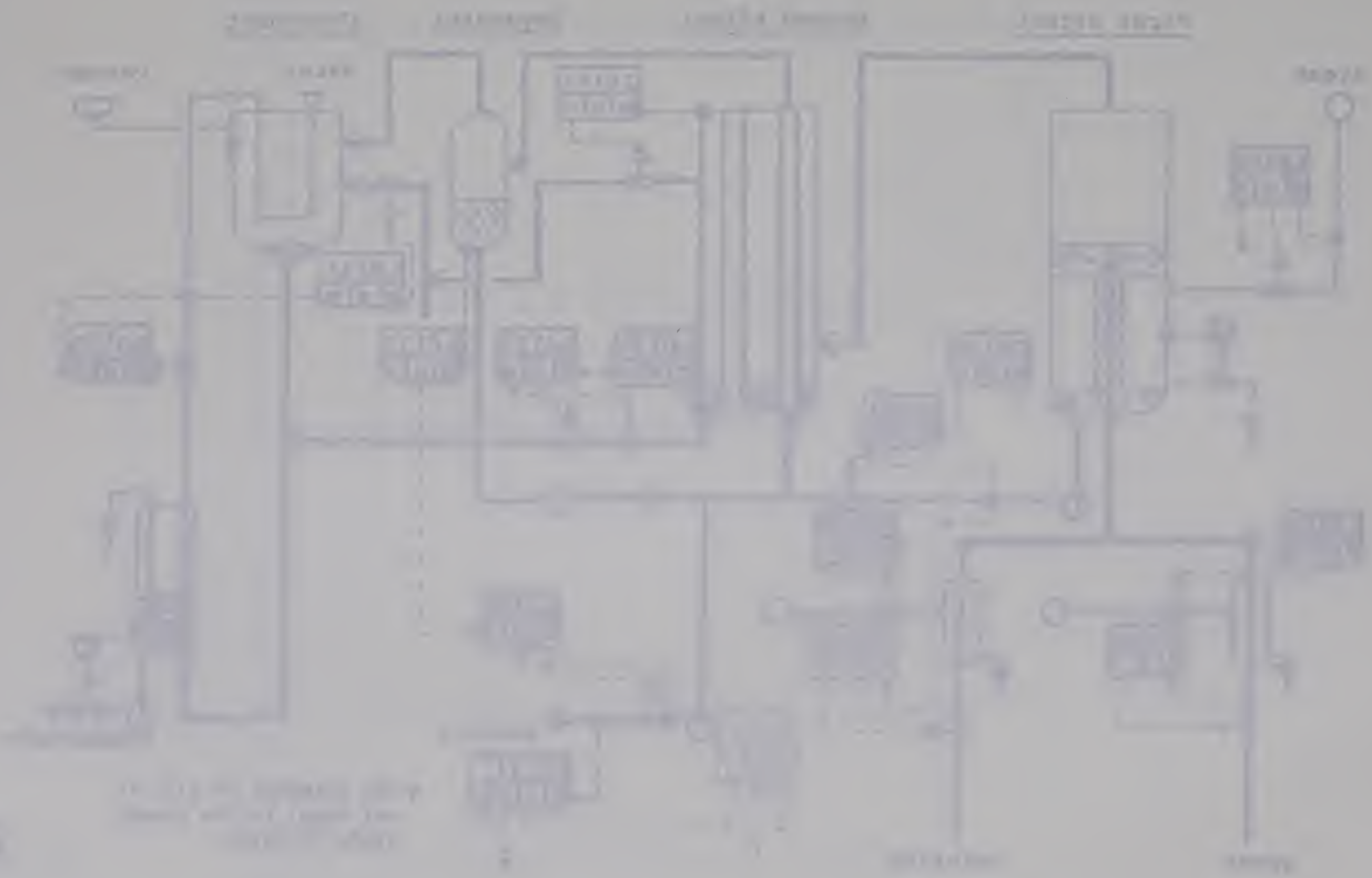


Table C-4

List of Orifice and Valve Stem Sizes

Process Variable Name	Orifice Size (inches)	Control Valve C_V
F1	0.382	4.0
F2	0.125	0.56
F5	0.100	0.28
F6	0.100	0.56
F7	0.150	--
F8	0.160	--
F9	0.327	2.25
F10	0.875	--
F11	0.125	0.20
F12	0.075	0.125
L12		1.13
L13		0.56
DVP21		0.012
P22		0.14
TT10		0.28
TT11		0.28

APPENDIX D

Evaporator Operating Procedures

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APPENDIX D

D.1 Introduction

Contained within this appendix are the various procedures used in the operation of the double effect evaporator. These include:

- startup procedure
- experimental run procedure
- procedure for unattended operation
- procedure for switching to and from direct digital control (DDC) and local analog control (LAC)

Reference should be made to the schematic drawings of the evaporator, Figures 2.1 and 2.2; and the detailed flow diagram (4) as an aid for the discussion.

Various physical operating configurations are available on the double effect evaporator (4) however, the discussion in this appendix will only be concerned with the forward feed mode shown in Figure 2.1. The total feed flow, consisting of a water and a water-glycol stream, enters the first effect after the two streams have been preheated and mixed to the desired feed temperature and concentration, respectively. Steam enters the first effect steam chest and is used to boil the liquid in this effect. The concentrated liquid solution is either pumped or flows to the second effect due to the pressure difference between the two effects. The second effect is operated under vacuum so

that the water vapor from the first effect can be used to heat the second effect. The liquid in this effect is continually cycled, under forced or natural circulation, from the separator through the heating tubes in the second effect steam chest back to the separator. The water vapor produced leaves the separator and is condensed and cooled in the condenser. This water is ultimately pumped back to the feed water tanks along with the condensate from the second effect steam chest. The concentrated liquid from the second effect is directed (using either the head from the circulation pump or the product pump) to the solution feed tanks.

D.2 Evaporator Startup Procedure

The following sections outline in detail the steps which must be performed in order to startup the double effect evaporator. It is assumed that:

- the operating mode of the evaporator is forward feed and the necessary hand values have been set to ensure this mode. (4)
- the evaporator will be started up using DDC.
- the operator is somewhat familiar with DDC and is familiar with "program BAJ30, that has been written to aid in the startup. (See Section D.2.4)

Once the operator is familiar with this procedure, other modes of operation including both the physical configuration and control schemes

can be implemented by following the basic steps in this procedure.

D.2.1 Prestartup Preparation

1. Turn all the current output stations to a manual mode and change the output of each so that the corresponding control valve is closed.
2. Turn on the main power to the evaporator by switching the main magnetic breaker on the side of the electrical panel.
3. Open the gate valves for each of the utilities at the corresponding overhead headers. These include:
   - air supply
   - vacuum supply
   - steam supply (V20)
   - cooling water supply (two valves)
4. Turn on the following solenoid switches:
   Number 2, 3, 5, 8, 12, 13, 14, 15, 16, 18. (See Table D-8 for description of solenoid valves). This ensures the following:
   - all the utilities, except the vacuum supply, are available to the evaporator
   - feed can be pumped, from the two water tanks and two solution tanks (E-10, E-11, E-2, E-5 respectively), to the first effect and recycled to each set of feed tanks.
   - product (F6) can be returned to the two solution tanks.
5. Close the gate valve (V66) at the discharge of the condensate

pump, E-26, to prevent air entering the vacuum system from the water feed tank, E-11.

6. Open the toggle valve (V33) on the first effect steam chest and the current output station (to 25.0%) for the first effect pressure control valve. This will allow air to be purged from both effects when steam enters the evaporator.
7. Fill the reference legs in each of the three level sight glasses (i.e., L14, L11, and L15) by turning on the water purge valves to the corresponding d-p cells.
8. Check that cooling water is going to:
   - the cooler on the packing gland seal of pump E-25.
   - the cooler (E-12) on the second effect condensate line (V55), which prevents the condensate from flashing.
   - the cooler (E-18), on the discharge line of pump E-26 (V46), which prevents the overheating of the condensate recycle stream.
   - the coolers (E8, E9) on the solution and water recycle lines of the evaporator feed system. (Open valves V15, V24, V18, V27).
9. Ensure that the necessary DDC loops (Table D-1 and D-2) are in the process variable table. If necessary load the deck of DDC control cards that define the standard evaporator control system (Figure 2.2). Reference the DDC Manual (7) for the procedure to follow.

Note: The evaporator can now be started up using the procedure in the next subsection. IF for some reason the operator cannot proceed to the startup, the evaporator can be left in the present state for an indefinite period of time with no serious consequences.

D.2.2 Startup Procedure

1. Turn on the following pumps:
   - feed water pump, E-21
   - feed solution pump, E-20
   - calibration pump, E-26 (this should only be turned on when the water level is visible in E-16)
   - rundown tank condensate pump, E-27 (it is suggested that this pump be bypassed during startup by bypassing the products from the condenser and second effect steam chest to E-16).

   Check the pressure gauges at each of the pump's discharge, to ensure that flow has been established. Note that flow has not been started to the first effect since the control valves on the feed flows, F11 and F12, are closed and therefore the water and solution feed is only recycling to the feed tanks.
2. Establish the liquid level (L14) in the first effect (E1) and liquid level (L11) in the separator (E13).

Note: If vacuum has not yet been applied, it is necessary to run pump E-23 to transfer fluid into the second effect; or manually close

V29, open V30 and put feed directly into the second effect.

3. Turn on the circulation pump E-25 and adjust the hand valve (V42) at the pump discharge until the corresponding flow (F10) is at 15.0 percent on the recording chart.

4. Change the controller constants in the first effect level (L14) and the separator level (L11) control loops so that each level is under "tight" control. The following constants, in hexadecimal notation, are suggested:

   - First Effect Liquid Level (DDC Loop 0136)

     Loop word 25 = 0000

     Loop word 26 = 0300

   - Separator Liquid Level (DDC Loop 0137)

     Loop word 25 = 0000

     Loop word 26 = 4FF0

   These changes must be made through the process operators consel (POC).

5. Change the evaporator DDC control loops, which are given in Table D-1 and D-2, to an operable manual mode. This can be done by queuing BAJ30 and entering option 1 (see Table D-5 for the options available in BAJ30).

Note: 1. BAJ30 uses free format input so that no specific input format is required. IF a number of values are to be entered in a given input set the user simply leaves a blank between each number entered. (Table D-5).

2. When the data has been entered, using the IBM 1816 typewriter, hit the "EOF" key to allow transfer of the data to the IBM 1800.
3. This procedure should be followed whenever BAJ30 is used.

6. Open solenoid switch number 1 to allow vacuum supply to the evaporator. Put the condenser vacuum (P22) loop, 0134, on automatic with a setpoint of -15.00 inches of Hg.

Note: 1. For those unfamiliar with DDC, single or cascaded DDC loops can be put to automatic using the procedure in D.2.4.
2. This procedure should be followed whenever a DDC loop is to be put on automatic.

7. Set the current output station on the cooling water flow, F9, to 85.0%. Put the corresponding DDC control loop, 0133, on automatic with a setpoint of +40.00 lbs/min. (reference D.2.4). Check pressure gauge PT22 to ensure that a vacuum is produced.
8. The next set of steps initiates flow to the first effect and establishes liquid level control on the first effect (L14) and separator (L11), respectively. These steps should be carried out without interruption.
   (a) Start feed flow into the first effect by:
      - Changing the setpoints of the water and solution feed DDC control loops, 0141 and 0139, to zero so

that the loops can be put on automatic with zero flow to the first effect. This can be done by queuing BAJ30 and entering option 5 (Table D-5) along with the following data for F, CF, XTK: 0.0 3.0 9.0 'EOF'

- Put 0139 and 0141 on automatic (reference Section D.2.4).
- Start feed flow into the evaporator at a flow rate of +1.000 lbs/min. To avoid diluting each effect with water, only allow the solution feed to enter. To accomplish this, queue BAJ30 and enter option 5. Enter F, CF, XTK as 1.0 1.0 1.0 'EOF'.

(b) Put the cascaded control system of first effect liquid level (L14) and first effect bottoms flow (F2) on automatic:

- Turn on the first effect bottoms pump, E-23 using a setting of 6.5 on the variable speed dial. Note that if desired, this pump can be left off providing the pump bypass valves have been properly set.
- Put the above cascaded control system on automatic I.E. DDC Loops 0.36 and 0151 (Reference Section D.2.4). Make sure the setpoint on 0136 is +22.00.

(c) Put the cascaded control system of the separator liquid level (L11) and the product flow rate (F6) on automatic:

- Turn on the product pump, E-24, using a setting of 7.5

on the variable speed dial. Check that the bypass valve back to the pump suction is partially opened. Note that if desired this pump can be left off providing the pump bypass valves have been properly set. If the latter configuration is used, it is suggested the total feedrate be increased from 1.0 lbs/min. to the desired feed flow in two steps. (see step 11)

- Put the above cascaded control system on automatic i.e. DDC loops 0137 and 0152 (Reference Section D.2.4). Make sure the setpoint on 0137 is +11.00.

9. Change the output on the steam's current output station to 3.0 percent. Also change the setpoint to +0.600 lbs./min. on the DDC loop, 0132 (i.e. queue BAJ30 and enter option 6. Enter the total feed flow rate as 1.5 lbs./min.). Note that during startup, the product concentration feedback controller, 0144, is left on manual.
10. When steams starts venting from the first effect steam chest, close the toggle valve (V33) on the chest. At this time the steam control loop (0132) should be put on automatic. (Reference Section D.2.4).
11. Increase the total feed rate and the steam flow rate to the desired final steady state. This can be done by queuing BAJ30 and entering option 7 (Table B-5). Enter the desired total feed flow and concentration (F and CF) and the actual solution

tank concentration (XTK). Note that the steam flow is set equal to 0.40 F.

12. Open the hand valve (V66) on the discharge of the condensate pump, E-26.
13. When the liquid level in the second effect steam chest (L15) starts to rise, put the cascaded system consisting of the second effect condensate level (L15) and the outlet condensate flow (F5) on automatic (i.e. DDC loops 0125 and 0150 respectively). Check that the setpoint in 0125 is at +08.00 inches of water. (Reference Section D.2.4).
14. Change the outputs to the two preheat exchanger's control valves to 20.0 percent through the COS (i.e. DDC loops 0138 and 0142). Allow a few minutes for the steam to purge the exchangers, then put the corresponding temperatures loops on automatic at a setpoint of +190.0 degrees F. (Reference D.2.4).
15. If it is desired to put the first effect under pressure control, put the DDC control loop (0145) on automatic with a set point of +02.00 inches of Hg. (Reference Section D.2.4).

Note: The evaporator startup has now been completed. The operator should allow it to come to steady state. Also follow the procedure in the next section.

D.2.3 Post Startup Procedure

1. Make a visual check of the evaporator to ensure proper operation of all the pumps, controllers etc.
2. Change the controller constants in the level loops, 0136 and 0137, back to their normal values after the evaporator has reached steady state.

D.2.4 Changing Single or Cascaded DDC Loops to Automatic Mode

This section is intended for those who are not familiar with DDC and must use BAJ30 to change a given DDC loop to an operable-automatic mode. The data required by BAJ30 for turning a single and cascaded control loop to this mode are given in Table D-1 and D-2, respectively. The output values given in each table are valid only on startup.

The operator should use the following steps:

1. Set the current output station (COS) output to the value given in Table D-1 or Table D-2. Note that the COS should be on manual.
2. Queue BAJ30 and enter the desired option (8 for single loops and 9 for cascaded loops to automatic). Enter the corresponding data from either Table D-1 or Table D-2 in free format (see Section D.2.2 step 5). Hit the EOF key on the IBM 1816 typewriter. (see Table D.9 for a typical input).
3. When a message has been written indicating the loop(s) are on automatic, put the COS to automatic.
4. Since the setpoint of the DDC control loop is set equal to the current measurement when the loop is put on the auto-

matic, it may have to be changed to a different value. In the startup procedure (Section D.2.2) this has been noted where applicable.

D.3 Experimental Run Procedure

The following description outline the steps that are required to initiate and complete a fully documented experimental run on the evaporator. This discussion assumes that the evaporator is running at the desired steady state with all the DDC control loops in the operable-automatic mode.

D.3.1 Prerun Preparation

Both the inline refractometers have a tendency to drift over a period of time, therefore prior to each run they should be zeroed using the following procedure:

1. Sample both the feed and product and place the test-tubes in the constant temperature bath.
2. After allowing sufficient time for the samples to cool, determine the concentration of each using the bench refractometer.
3. Adjust the inline refractometers by turning the zero adjust knob until the chart reading corresponds to the converted concentration reading i.e.

$$\text{For CF Chart Reading (\%)} = \frac{\text{BRR(\%)}*100.00}{6.0}$$

$$\text{For C2 Chart Reading (\%)} = \frac{\text{BRR(\%)}*100.0}{19.1}$$

where BRR = bench refractometer concentration reading. (Note: ensure the product concentration controller is on manual).

4. Repeat the procedure every 15 minutes until no adjustment is necessary.

D.3.2 Initial Steady State Documentation

Two Process and one Nonprocess program must be run in order to obtain a printout of the steady state data. These are:

1. MEBOL - This program computes the errors of closure in the mass, component and energy balances around the evaporator. The program execution is initiated by queuing the process program MEBOL. A typical input, via the IBM 1816 typewriters, is shown in Table D-2.

2. PVLOG - This program prints the current measurements of the various process variables such as flow rates, temperatures, pressures etc. on the IBM 1132 line printer. The program execution is initiated by queuing the process program PVLOG.

3. BJ20D - This program reads and prints the measurements of all the evaporator temperature points that have been interfaced with the IBM 1800. Execution is initiated through the Nonprocess monitor. A listing of the necessary input cards is given in Table D-4.

D.3.3 Control Program Initialization

A detailed discussion of the input data required is given in Ref. (24). To initiate the control program simply place the data cards in the IBM 1442 card reader and queue the process program BAJ10. Once initialized the control program is automatically queued every control interval and:

- does data acquisition and filing of the specified variables at the time interval specified by the user.
- executes control algorithms in conjunction with or in place of DDC.

If any DDC data accumulation loops are being used to buffer data to disk, the DDC file counter should be initialized at this time. This can be accomplished by queuing BAJ30 and entering option 4. A typical input for this process program is shown in Table D-5.

D.3.4 Load Disturbance Initiation

Before the disturbance is entered, allow approximately 10 minutes for initial steady state data accumulation. All disturbances can be initiated through the process program BAJ30. These include:

- Feed Flow Rate
- Feed Concentration
- Feed Temperature

As shown in Table D-5 these are options 5, 5 and 10 respectively. For the feed flow and/or concentration disturbances specify the desired flow rate, desired composition and the composition of the solution feed tank.

For the temperature disturbance just enter the desired preheat temperature. The changes are implemented immediately by sending a new setpoint to the DDC loops.

D.3.5 Concentration Monitoring

As mentioned in B.3.1, the inline refractometers are not completely reliable, therefore a periodic checking procedure is required to ensure they are functioning properly. This involves taking samples during the run and comparing the concentration of these samples with the inline refractometers. If there is a significant difference, rezero the instrument to the correct value by adjusting the instrument's zero.

D.3.6 Run Termination

When the run has been completed, duplicate samples of both the feed and product should be taken to establish the final steady state concentrations. Also a printout of the final steady state documentation should be obtained following the procedure of D.3.2.

The transient data are recovered by using the punching and plotting routines, BJ20A and BJ20B respectively, which are run as Non-process programs. To punch the data, the user must use the appropriate card deck to initiate the operation and put blank cards in the card reader. A message will be written on the IBM 1816 typewriter instructing the user to turn on digit switch zero. Before plotting, ensure that the plotter is ready to use and then enter the cards in the card reader. If more than one plot is being made, the user has control of when they will

be plotted by setting digit switch zero:

- up position program plots
- down position program cycles

It should be noted that plotting can be done using the punched data cards therefore the plotting can be carried during the next run. A detailed user description of both these programs is given in Ref. (24).

If data was buffered to disk through DDC, the transient data can be plotted using the Nonprocess program THIS. The necessary data is inputted via the IBM 1816 typewriters. A typical input is shown in Table B-6. This data must be recovered before the DDC file counter has been reinitialized.

D.4 Unattended Operation

The evaporator can be left unattended by using a FORTRAN written program to monitor the key variables. The program is executed at a preset sampling interval and checks the particular variable against a high and low limit. If the particular limits have been exceeded, for only one of the variables, the evaporator is completely shutdown in a fail safe condition. This is accomplished by activating one electronic contact which turns off the main power supply. (See Ref. (24) for a more detailed discussion). If the evaporator is shutdown, the operator should startup or completely shutdown by closing the hand valves for each of the utilities at the corresponding overhead header.

In order to prepare the evaporator for unattended operation the following is suggested:

- set the feed flow rate and the steam flow rate at 4.5 and 1.8 lbs./min., respectively.
- put the product concentration controller on manual control
- if possible, turn off the three Jabsco pumps (E-27, E-23, E-24) and adjust the hand valves for this configuration.
- if desired turn off the Foxboro charts.

When the evaporator has reached steady state turn on the monitor program by queuing BAJ40 and entering the desired option. Enter either the

1 = Start monitor program

3 = Reinitialize the data file in BAJ40 to the current steady condition and start the monitor program.

The program can be stopped by queuing BAJ40 and entering 2.

D.5 Switching - LAC/DDC

Currently the evaporator can be controlled using either direct digital control or local analog control. Switching between these modes has become necessary due to planned shutdowns of the IBM 1800. The procedures in the next subsections outline the steps to follow in order to obtain a bumpless transfer between the two control modes. Table D-7 outlines the associated DDC and LAC controllers. The evaporator should be at steady state before switching between the two modes.

D.5.1 Switching from DDC to LAC

To obtain a bumpless transfer from DDC to LAC the following steps should be followed for each control loop:

1. Change the LA controller to manual control.

2. Adjust the manual control knob on the local controller until the output meters on the current output station and the corresponding LA controller agree.
3. Switch the control to the LA control by pressing the LAO button on the interface module.
4. A visual check of the recording chart, for the variables which are recorded, will indicate if this phase was accomplished bumplessly. Small adjustments of the manual control knob can be made at this time.
5. Move the transfer switch on the LA controller to the manual-balance position. Adjust the setpoint dial until the deviation indicator on the LA controller is zero. (The deviation indicator registers the difference between the manual output and the automatic controller output when the transfer switch is in the manual-balance position).
6. Move the transfer switch to automatic.

D.5.2 Switching from LAC to DDC

To change the mode of control from LAC to DDC, perform the following steps for each control loop:

1. Put the corresponding DDC loop(s) in an operable - manual mode.
2. Ensure the current output station is on manual.
3. Adjust the COS output until the output meters on the COS and the LA controller agree.
4. Switch control to COS manual by pressing the LAO button.

5. Follow the procedure outlined in Section D.2.4, omitting the first step.

Table D-1

BAJ30 Data for a Single DDC Loop to Automatic

DDC Loop Description	Loop ID Number (ID)	Output in Percent (OUT)	MSSS Units Code (JDM)
Condenser vacuum pressure, P22	0134	5.0	6
Solution feed flow, F12	0139	0.0	4
Water feed flow, F11	0141	0.0	4
Cooling water flow, F9	0133	85.0	4
Steam flow, F1	0132	3.0	4
Solution feed temperature, TT10	0138	85.0	2
Water feed temperature, TT11	0142	85.0	2
First effect pressure, P20	0145	25.0	6

Table D-2

BAJ30 Data for Cascaded DDC Loops to Automatic

DDC Loop Description	Outer Loop ID (IDM)	MSSS Units Code for IDM (JDM)	Inner Loop ID (IDS)	MSSS Units Code for IDS (JDS)	CONA Eng.Units (CONA)	CONB Eng.Units (CONB)	Output in Percent (OUT)
First effect liquid level first effect bottoms flow (L14 - F2)	0136	5	0151	4	12.48	0.0	60.0
Separator liquid level product flow (L11-F6)	0137	5	0152	4	5.34	0.0	50.0
Second effect condensate level - flow (L15 - F5)	0125	5	0150	4	6.46	0.0	40.0

Table D-3

MEBOL Input Data

```
ENTER NAME PRIORITY ERROR
MEBOL
  MEBOL HAS BEEN QUEUED
                                  MAR69 MEBOL
     INPUT MODE NUMBER PLUS DESIRED OPTIONS
     J,K,L J IS MODE NO AS PER OPERATING SPECS
     J EQUALS ZERO STOPS EXECUTION
     K IS THE DESIRED SCAN TIME FOR DATA ACQUISITION
     L IS THE NUMBER OF MEASUREMENTS REQUIRED IN AVERAGING
       TECHNIQUE

  4 4 5  (Hit "EOF")

     INPUT LOGICAL UNIT NUMBERS FOR
     OUTPUT OF PROGRAM OPERATOR COMMUNICATIONS
     OUTPUT OF RESULTS

  4  6   (Hit "EOF")

                              DATA ACQUISITION STARTED


          ENTER OPTION
         -1  STOPS EXECUTION AFTER BALANCE  CALCULATIONS
          0  STOPS EXECUTION IMMEDIATELY
          1  SAME AS -1 PLUS DATA ADJUSTMENT

-1   (Hit "EOF")

     ENTER AMBIENT ROOM TEMPERATURE

75.0 (Hit "EOF")
```

Table D-4

BJ20D Input Data

```
       4      10     15     20     40
//     JOB     X      X      X     USER NAME

       4     8      16
//     XEQ   BJ20D  FX
```

DATA CARDS WITH FOLLOWING FORMAT

```
I5                          25A2
(MULTIPLEXER ADDRESS)       (DESCRIPTION)

//  JOB     X     X     X

//  END

BLANK CARD
```

Table D-5

BAJ30 Input Data

```
ENTER NAME PRIORITY ERROR
BAJ30
 BAJ30 HAS BEEN QUEUED

  ENTER 1 = NEED HELP  2=CONTINUE
1

  ENTER  1=START LOOPS(OP,MN)  2=START LOOPS(OP,AU)
         3=STOP LOOPS  4=INITIALIZE DISK TRANSFER
         5=CHANGE F OR CF  6=CHANGE S  7=BOTH 5 AND6
         8=SINGLE LOOP AUTO  9=CASCADED LOOP AUTO
         10=CHANGE TF
5
      ENTER F,CF AND CTK RESPECTIVELY
5.50 3.00 9.15 (hit "EOF")
```

Table D-6

<u>THIS Input Data</u>

```
    ENTER ID OF DATA ACQUISITION LOOP, POLL TIME INTERVAL IN SECONDS
0297 32
    DATA TRANSFER TO TEMPORARY FILE COMPLETED
    NUMBER OF POINTS TRANSFERRED IS 1500
     ENTER NO. OF POINTS TO BE USED
100
    ENTER TIME UNITS  1=SEC. 2=MIN. 3=HRS.
2
    TOTAL TIME TO APPEAR ON ABSCISSA IS    53.3
    ENTER INFORMATION ON GRAPH IN THIS ORDER
    MAXIMUM VALUE OF DEPENDENT VARIABLE(REAL)
    NUMBER OF DIVISIONS ALONG ABSCISSA(INTEGER)
    NUMBER OF DIVISIONS ALONG ORDINATE(INTEGER)
300.0 100.0 4 4
    ENTER THE LABELS FOR THE Y AND X AXIS RESPECTIVELY
    ENTER ONE AT A TIME . MAX. OF 30 CHARACTERS EACH
TEMPERATURE (DEG. F.)

TIME  (MINUTES)
      ENTER TITLE USING MAX. OF 30 CHARACTERS
TEST PLOT
```

Table D-7

DDC and LAC Control Loops

Identification Number for DDC Control Loop	Process Variable Name for Local Analog Control Loop
0139	FC12
0141	FC11
0138	TC10
0142	TC12
0144-0132	FRC1
0136-0151	LRC14
0137-0152	LRC11
0125-0150	LC15
0133	FRC9
0134	PRC22
0145	Leave on COS manual

Table D-8

Description of Solenoid Switches

Solenoid Switch Number	Description of Location
1	Vacuum supply header
2	Water feed flow to first effect
3	Cooling water flow at condenser inlet
4	Feedwater recycle line to solution tank, E-2
5	Feed solution recycle line to solution tank, E-5
6	Feed water recycle line to solution tank, E-5
7	Feed solution recycle line to water tank, E-10
8	Feed water recycle line to water tank, E-11
9	Water recycle line to solution tanks, E-2, E-5
10	Bypass around control valve DVPC21
11	Main power supply
12	Solution feed flow to the first effect
13	Feed solution recycle line to solution tank, E-2
14	Steam supply header
15	Air supply header
16	Product line to solution tank, E-2
17	Product line to water tank, E-10
18	Product line to solution tank, E-5

Table D-9

BAJ30 Input Data

```
ENTER NAME PRIORITY ERROR
BAJ30
 BAJ30 HAS BEEN QUEUED

  ENTER 1= NEED HELP  2=CONTINUE
1

  ENTER  1=START LOOPS(OP,MN)  2=START LOOPS(OP,AU)
         3=STOP LOOPS  4=INITIALIZE DISK TRANSFER
         5=CHANGE F OR CF  6=CHANGE S  7=BOTH 5 AND 6
         8=SINGLE LOOP AUTO  9=CASCADED LOOP AUTO
         10=CHANGE TF
8
      ENTER ID,OUT,JDM
0145 30.0 6 (hit "EOF")
      LOOP   145  AUTO
```